

Economic Evaluation of an Advanced Super Critical Oxy-Coal Power Plant with CO₂ Capture

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AUTHOR'S DECLARATION

I hereby declare that I am the sole author of this thesis. This is a true copy of the thesis, including any required final revisions, as accepted by the examiners.

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ABSTRACT

Today's carbon constrained world with its increasing demand for cheap energy and a fossil fuel intensive fleet of power producers is making carbon capture and storage (CCS) desirable. Several CCS technologies are under investigation by various research and development groups globally. One of the more promising technologies is oxy-fuel combustion, since it produces a CO₂ rich flue gas which requires minor processing to meet storage condition requirements. In this study the economics of an advanced super critical oxy-coal power plant burning lignite, simulated in-house was assessed. A robust and user-friendly financial tool box has been developed with commonly acceptable default parameter settings. Capital, operation and maintenance costs were estimated along with corresponding levelized cost of electricity and CO₂ avoidance costs calculated using the detailed financial model developed. A levelized cost of electricity of 131 \$/MWhr_{net} along with a levelized CO₂ avoidance cost of 64 \$/tonne was estimated for an ASC oxy-coal power plant with CO₂ capture. Also a levelized cost of electricity of 83 \$/MWhr_{net} was estimated for an ASC air-fired coal power plant without CO₂ capture capabilities as the base plant. The price of electricity was observed to increase from 83 \$/MWhr_{net} to 131 \$/MWhr_{net} translating into a 57% increase. The sensitivity of the overall economics of the process was assessed to several parameters. The overall economics was found sensitive to the choice chemical engineering plant cost index (CEPCI), capacity factor, size of power plant, debt ratio, fuel price, interest rate, and construction duration.

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DEDICATION

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Table of Contents

LIST OF TABLES	x
LIST OF FIGURES	xi
1 INTRODUCTION	1
2 LITERATURE REVIEW	2
2.1 The approach	2
2.2 Oxy-fuel combustion process components	7
2.2.1 Coal preparation & ash handling	9
2.2.2 Air separation unit.....	12
2.2.3 Boiler, steam cycles and flue gas recycle	17
2.2.4 Flue gas clean up.....	23
2.2.5 Flue gas compression	28
2.3 Degree of integration.....	31
3 COST COMPARISON OF CARBON CAPTURE TECHNOLOGIES.....	34
3.1 Coal power generation technologies	35
3.1.1 Conventional Coal Combustion with CO ₂ capture	35
3.1.2 Gasification process	37
3.1.3 Oxy-fuel combustion with CO ₂ capture.....	42
3.2 Multi-technology assessment studies	44
4 ECONOMICS	48
4.1 Financial modeling.....	48
4.1.1 Capacity factor	48
4.1.2 Fuel consumption.....	49
4.1.3 Escalation parameters	49
4.1.4 Financial profile	49
4.1.5 Construction duration.....	49
4.1.6 Taxation	50
4.1.7 Depreciation and capital cost allowance.....	50
4.1.8 Distribution of investment requirements during the construction period	50
4.1.9 Levelized cost of electricity	51

4.1.10	Cost of CO ₂ avoided	51
4.1.11	Summary of assumptions	52
4.2	Cost estimation methodology	53
4.2.1	Capital cost estimation	53
4.3	Capital cost estimation results	57
4.3.1	Plant description	58
4.3.2	Coal/sorbent preparation, ash & spent sorbent handling	60
4.3.3	Air separation unit	61
4.3.4	Boiler and accessories	62
4.3.5	Balance of plant (BOP)	63
4.3.6	Flue gas clean up	64
4.3.7	CO ₂ capture and compression unit	65
4.3.8	Utilities and off sites	66
4.3.9	Summary of capital costs and cost estimation methodology	67
4.4	Operations & maintenance costs (O&M)	70
4.4.1	Fixed costs	70
4.4.2	Variable costs	71
4.4.3	Comparison of O&M costs	73
4.5	Summary of costs	75
5	DESIGN CONSIDERATIONS	76
5.1	CO ₂ compression unit design considerations	76
5.1.1	Materials of construction	76
5.1.2	Heat exchanger design	76
5.1.3	Compressor design	77
5.2	Flue gas clean up	78
5.2.1	Materials of construction	78
5.2.2	Heat exchanger design	78
5.2.3	Overdesign factor	79
6	SENSITIVITY ANALYSIS	80
6.1	Chemical engineering plant cost index	80

6.2	Fuel price.....	81
6.3	Capacity factor	82
6.4	Interest rate	83
6.5	Debt ratio.....	84
6.6	Construction duration.....	85
6.7	Economic life of plant	86
6.8	Power plant size	87
6.9	Sensitivity analysis summary	88
7	CONCLUSIONS & RECOMMENDATIONS.....	91
	REFERENCES	94
	APPENDICES	99
	Appendix 1: vendor quotations	99
	Aluminum brazed heat exchangers	99
	Molecular sieve CO ₂ dryer	104

LIST OF TABLES

Table 1: Types of capital cost estimates [37].....	5
Table 2: Summary of heat integration processes [14]	32
Table 3: Split heating sources required for the drying of coal [19]	32
Table 4: Common gasifier types and their corresponding characteristics [8]	40
Table 5: Environmental contrasts between a combustion and a gasification process [15]....	41
Table 6: Cost comparison of power generation technologies with CO ₂ capture (costs in 2008 USD)	46
Table 7. Economic assumptions	52
Table 8: Cost estimation methodologies employed for the ASC oxy-coal with CO ₂ capture	54
Table 9: Summary of capital costs	69
Table 10: Maintenance cost summary	71
Table 11: Operations and maintenance cost summary	72
Table 12: Summary of costs	75

LIST OF FIGURES

Figure 1: CO ₂ capture pathways [22].....	2
Figure 2: New technology deployment curve for coal [9].....	4
Figure 3: Oxy-fuel CO ₂ capture plant concept burning lignite [46]	8
Figure 4: MEA Absorption [35]	36
Figure 5: Gasification phase diagram [15].....	37
Figure 6: IGCC process with CO ₂ scrubbing [35]	39
Figure 7: Gasification types by technology [15].....	41
Figure 8: IGCC availability history (excluding operation on back up fuel) [18]	42
Figure 9: Oxy-fuel combustion [35]	44
Figure 10: ASC oxyfuel combustion with CO ₂ capture [48].....	58
Figure 11: Three column design cryogenic air separation unit.[2].....	59
Figure 12: CO ₂ capture and compression unit [47]	60
Figure 13: Coal/sorbent preparation, ash & spent sorbent handling costs for a 380.5 MW _{net} ASC oxy-coal plant with CO ₂ capture	61
Figure 14: Air separation unit's capital costs for a 380.5 MW _{net} ASC oxy-coal plant with CO ₂ capture.....	62
Figure 15: Boiler & accessories' capital costs for a 380.5 MW _{net} ASC oxy-coal plant with CO ₂ capture.....	63
Figure 16: Balance of plant's capital costs for a 380.5 MW _{net} ASC oxy-coal plant with CO ₂ capture.....	64
Figure 17: Flue gas cleanup unit's capital costs for a 380.5 MW _{net} ASC oxy-coal plant with CO ₂ capture.....	65
Figure 18: Capture and compression unit's capital costs for a 380.5 MW _{net} ASC oxy-coal plant with CO ₂ capture.....	66
Figure 19: Utilities and off sites' capital costs for a 380.5 MW _{net} ASC oxy-coal plant with CO ₂ capture.....	67
Figure 20: Summary of capital costs for a 380.5 MW _{net} ASC oxy-coal plant with CO ₂ capture.....	68
Figure 21: Comparison of O&M costs for a 380.5 MW _{net} ASC oxy-coal plant with CO ₂ capture.....	74
Figure 22: Sensitivity of LCOE with and without capture & LACOC to CEPCI.....	81
Figure 23: Sensitivity of LCOE with and without capture & LACOC to fuel price	82

Figure 24: Sensitivity of LCOE with and without capture & LACOC to capacity factor.....	83
Figure 25: Sensitivity of LCOE with and without capture & LACOC to interest rate.....	84
Figure 26: Sensitivity of LCOE with and without capture & LACOC to debt ratio	85
Figure 27: Sensitivity of LCOE with and without capture & LACOC to construction duration	86
Figure 28: Sensitivity of LCOE with and without capture & LACOC to economic life of plant.....	87
Figure 29: Sensitivity of LCOE with and without capture & LACOC to plant size	88
Figure 30: Summary of sensitivities of LCOE with CO ₂ capture to parameters of consideration	89
Figure 31: Summary of sensitivities of LACOC to parameters of consideration.....	90

1 INTRODUCTION

Rapid growth in energy demands along with emerging carbon constraints have forced the research and development sector to investigate green and sustainable energy production technologies in the past decade. A significant source of anthropogenic greenhouse gas emissions is the power generation industry. Power generation plants are large single point emitters, thus capturing the emissions from these plants is considered to be an effective solution to achieve significant reductions in greenhouse gas emissions. It should be noted that fossil fuels play an important role as energy source in the energy production industry both globally and in North America. Amongst the fossil fuels the cheapest and most abundant of all is coal. It has one of the highest emission intensities of all fossil fuels; however, it cannot be rapidly eliminated from the power production fuels' selection mix since its hasty elimination will cause disturbances in the reliability of energy supply. Thus carbon capture and storage is believed to be a reliable midterm solution for greenhouse gas mitigation from power plants while meeting the market's energy demands reliably.

2 LITERATURE REVIEW

2.1 The approach

Three non-pressurized carbon capture methods are currently being investigated by various energy sectors and academia internationally in search of finding the most economical clean coal energy production option. They are:

- post combustion CO₂ capture using solvent absorption;
- pre-combustion capture using various types of integrated gasification combined cycles with CO₂ capture; and,
- oxy-fuel combustion with CO₂ capture.

A simplistic flow sheet of the three processes is provided in Figure 1.

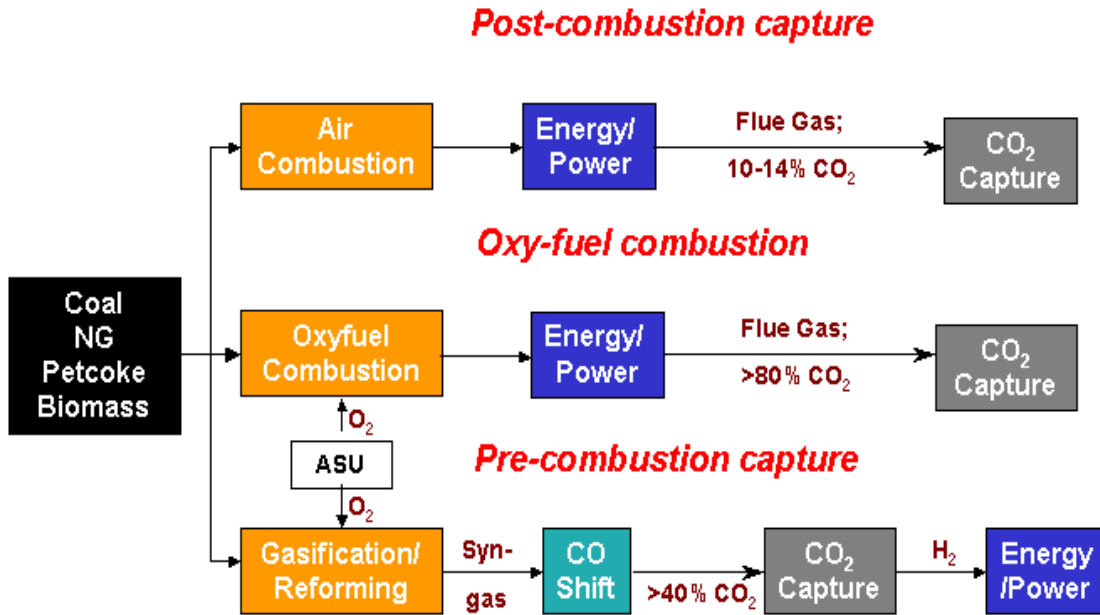


Figure 1: CO₂ capture pathways [22]

The task of economic evaluation of these processes is a cumbersome one due to their high complexities. As of today, there is no definitive answer as to which technology is superior *a*

priori to a detailed specification/analysis of fuel, economic variables and plant characteristics. The high variability observed in the economics presented for each technology can be attributed to the effects of:

- type of coal;
- various process configurations possible;
- choice of equipment/technology used for the individual comprising processes;
- degree of process integration minimizing parasitic energy loss;
- footprint limitations;
- price of fuel and chemicals;
- economic parameters;
- CO₂ tax/credit system in place;
- Location; and,
- maturity of the technology.

From the above mentioned factors, the effect of maturity of the technology cannot be evaluated in a quantitative manner; however readers and decision makers should take it into consideration before drawing conclusions from techno-economic analyses' results presented to them. A recent report by the Canadian Clean Power Coalition makes note of this issue and presents in Figure 2 a common trajectory and experience curve for technical developments applied to power production technologies employing coal as the fuel.

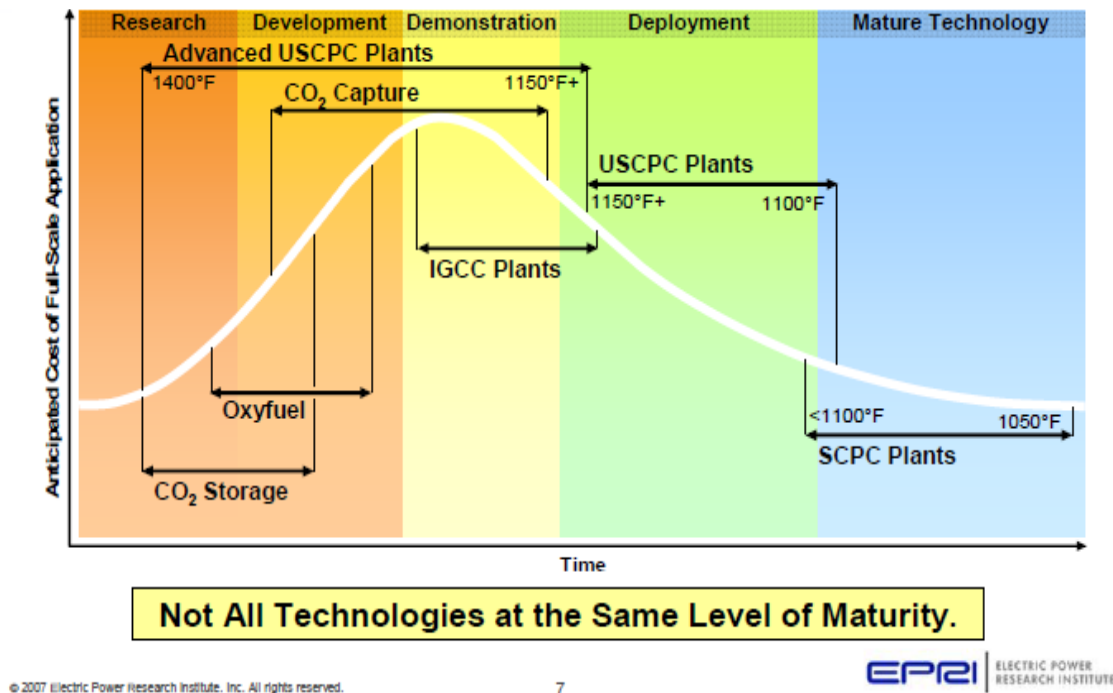


Figure 2: New technology deployment curve for coal [9]

The common path starts with a promising technology being proposed. As the technology is better understood, and the possible issues are identified the projected costs start to escalate until the near commercial demonstration stage at which time most of the possible issues and the associated risks are evaluated. After a successful demonstration at near commercial scale where the major benefits are confirmed and all related issues identified the stage of incremental improvement starts until the process achieves full maturity and shows its maximum potential. There are no shortcuts to this trajectory, thus the best way to evaluate a relatively immature technology, is to prepare detailed simulations with detailed site specifications to identify possible issues that might be of concern [9]. However it should be noted that for the case of oxy-coal combustion due to its nature of utilizing known technologies and an already existing supply chain it is believed that the technology will travel through the learning curve and attain maturity for a commercial demonstration level fairly quickly [30].

As for the detailed simulations of power plants to determine a suitable technology for a certain set of conditions, even after site specifications are made available, large amount of time and resources will be required to rigorously model, simulate and optimize all the scenarios while considering all possible permutations of processes and configurations combinations. The approach taken by some studies to overcome this issue has been to study all the promising options at a preliminary level initially and develop more accurate models for the ones shown to be the most competitive. However this approach might be problematic as will be explained shortly.

It should be noted that the level of accuracy of techno-economic studies are commonly categorized in the following manner and the proposed nomenclature is employed throughout the remainder of this report:

Table 1: Types of capital cost estimates [37]

Type	Description	Accuracy
Order-of-magnitude	Based on similar previous cost data	$>\pm 30\%$
Study	Based on knowledge of major items of equipment	$\pm 30\%$
Preliminary	Based on sufficient data to permit estimate to be budgeted	$\pm 20\%$
Definitive	Based on almost complete data but before completion of drawings & specifications	$\pm 10\%$
Detailed	Based on complete engineering drawings specifications, and site surveys	$\pm 5\%$

A comprehensive report assessing low rank coal power plants [19] has approached the issue by first carrying out an order-of-magnitude assessment for all options and then selecting the more promising ones for a preliminary analysis to determine the most suitable choice of technology, as previously proposed. Although this approach might seem logical, due to the presence of large confidence intervals around the estimations and the competitive state of the technologies, these initial assessments' results might only indicate differences well within/smaller than the error of the actual studies. For instance the initial assessment results of the aforementioned study illustrated that all case scenarios were within 10% of each other

in terms of COE (cost of electricity), while the individual studies' accuracies were to be in the order of $\pm 35\%$. Thus selecting the best two choices is potentially an erroneous conclusion from a statistical point of view and the optimal result chosen has a high probability to be in fact a non-optimal option. Other studies have only focused on one technology and considered with and without capture scenarios and have evaluated values such as cost of capture, de-rate (reduction in electrical power rating of the power plant) imposed on the base plant efficiency (without capture), and the amount of CO₂ avoided [3][14][16][40]. Other studies have further identified a break even cost of CO₂ (BE). The break even cost of CO₂ is determined by driving up the credit price for CO₂ until the first year cost of power for a reference coal plant without capture equals the first year cost of power for the carbon capturing plant [9].

Based on the various approaches observed in the literature it can be seen that a reasonable approach would be to choose a technology that seems practical based on some of the aforementioned critical factors and analyze it as rigorously as reasonably possible, both without capture and/or at a capture ready level and finally with full capture capabilities. Due to the large debate on the definition of capture ready plants a detailed definition of the terminology "capture ready" is recommended to be provided *a priori* to the start of a study. A report from IEA [25] assesses this issue and recommends a set of requirements that need to be considered in order to be able to claim that a plant is 'capture ready'. Furthermore several studies have assessed various configuration possibilities specific to a certain technology to evaluate their feasibility and cost reduction opportunities [3][16].

2.2 Oxy-fuel combustion process components

Due to the highly competitive state of the three main technologies mentioned, any gain in efficiency could result in the superiority of one alternative over another. Furthermore, studies have found oxy-fuel combustion to be a high efficiency technology, yet a relatively capital intensive choice for low rank coals [9][19]. Thus identification of alternatives, for the capital and energy intensive individual processes, has been one of the main considerations of many researchers of this technology.

The main difference between oxy-fuel combustion and air-fired combustion is the major reduction in the amount of inert gases in the combustion gas for oxy-fuel combustion which needs to be addressed if similar combustion characteristics are desired. In oxy-fuel combustion the combustion air is separated before entering the boiler removing its 79% nitrogen, and most other impurities, resulting in a purified oxygen stream. The purified oxygen stream is fed to the boiler along with a large portion of the flue gases which is recycled to make up for the nitrogen that is removed from the combustion gas. For both greenfield power plants and retrofit air-fired boilers this recycle stream needs to be put in place in order to keep mass flow rates similar to those of air-fired combustion. This is so that the convection heat transfer of the boiler would be similar, and the flame temperatures are kept at low enough levels. To obtain similar combustion characteristics at least 2/3 of the flue gas needs to be recycled as stated by some sources while others propose a minimum of about 70% with the remaining being pure oxygen [3][14][16][19][40]. The aforementioned recycle streams can be extracted at various points along the flue gas cleanup process. Typically a two stream configuration is proposed. A primary recycle stream being the cleaner and cooler of the two, is used for carrying the fuel. A secondary recycle stream, typically hotter and wetter than the primary stream is used for carrying oxygen and increasing combustion gas flow rates to achieve similar combustion characteristics to the air-fired cases. A schematic diagram of concept oxy-fuel power plant burning lignite, extracted from a recent study performed by the BERR [46], can be seen in Figure 3.

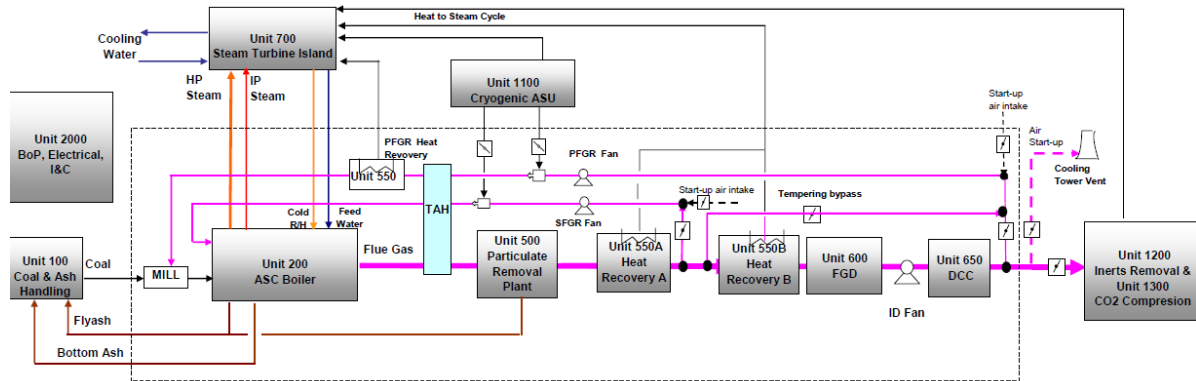


Figure 3: Oxy-fuel CO₂ capture plant concept burning lignite [46]

Similarly in this study an oxy-fuel plant is sectioned, having the following process areas which are further discussed in the proceeding sections.

- Coal preparation & ash handling;
- Air separation;
- Boiler, steam cycles and flue gas recycle;
- Flue gas clean up; and
- Flue gas compression.

2.2.1 Coal preparation & ash handling

2.2.1.1 Moisture handling

Coal arriving from the coal mine comes in large pieces which contain high levels of ash and moisture (dependent on the type of coal). Thus, typically, coal needs to be crushed, subsequently pulverized and finally fluidized with combustion air or gas prior to entering the boiler. Significant efficiency gains, however, have been proposed from utilizing the low temperature heat available in the plant (that would normally be wasted) to dry coal in order to improve the combustion characteristics. In pulverized fuel combustion with air this task can be fulfilled by heating the combustion air stream upstream from the boiler and employ it as the pulverizing fluid. As for the case of oxy-fuel combustion this task is not so simple due to the fact that the intuitive replacement stream for air, the combustion gas stream, is pure oxygen, and due to highly flammable environments produced when oxygen is present in this stream, it is not a suitable choice. The alternative replacement for the air stream thus becomes the flue gas recycle stream prior to mixing with the pure oxygen stream. However, the impurities such as acid gases present in the recycle flue gases could potentially cause various problems that need to be addressed. In most recent literature the drying process has been an unavoidable component for the case of oxy-fuel combustion for all coal types due to its associated efficiency gains [14][19][21][26].

An IEA study [19] expressed concerns about the drying temperature, in particular for direct drying of coal with the recycle flue gas stream. It notes that certain volatiles, which have significant effects on good ignition and combustion characteristics, will be lost if the drying temperature is too high. At temperatures above 120°C CO₂ is liberated, and at temperatures above 180°C H₂, CO and CH₄ evaporate from coal, but the study still promotes coal drying and associates the following advantages to its usage:

- An efficiency gain of a few percentage points in the overall efficiency;
- Decrease in mill power consumption by 6% due to reduced flows; and,

- Reduction in flue gas flow rate by 6%, thus reducing the size of flue gas treatment processes downstream.

There are different ways that this low temperature drying task can be fulfilled. The primary recycle by many authors have been chosen for this job. An IEA study [14] proposes drying of coal by nitrogen/air taken from the ASU (Air Separation Unit). This involves an indirect heating system, in which the hot nitrogen/air stream carries the coal to the mill and, after milling, they are separated and the coal is reintroduced to the recycle flue gas stream. However the study avoided using this coal drying technique due to its large deviation from robust known milling operations currently being used commercially.

Another IEA study [19] suggests that drying coal with the hot nitrogen stream is more promising. This study claims that the energy consuming pressure drop caused by the direct contact with combustion flue gas, along with the significant anticipated amounts of coal fine entrainment and the cumbersome task of controlling the drying temperature are justifications for its proposal. In this process after drying the flue gas the primary recycle stream is used to fluidize the coal, downstream of the milling operation into the boiler. The fluid carrying the moisture away is then sent to a bag filter before being vented out to the atmosphere.

Another comprehensive study by Canmet Energy [21] has been devoted strictly to assessing the removal of moisture from coal using the high purity nitrogen produced from the ASU and its effects on the boiler efficiency, oxygen requirements, emissions, overall plant net efficiency and cooling water requirements. The study proposes using the cold dry nitrogen for cooling purposes in the CO₂ compression train initially. The heated dry nitrogen stream is then employed for the drying of coal. Furthermore, the study illustrates that the drying of the lignite from 33% to 10% moisture content decreases the required coal input by 32%, increases the efficiency of the boiler by 5%, and decreases the oxygen demand by 6%. Decreasing the moisture content also results in reduction in size of all downstream pollution control equipments as well as that of the boiler. There is also a reduction in compressor cooling load requirements due to increased efficiency, along with a reduction in cooling

water requirements for the same power output. The amount of flue gas to be treated is reduced by 8% resulting in the lowering of CO₂ transportation and storage costs. Furthermore the calculated overall net plant efficiency increase is reported to be 2.5 percentage points. The study also assesses the flow requirements of nitrogen and air for the drying of coal to three different moisture levels. It claims that the drying tasks can be fulfilled using a combination of dry nitrogen, ambient air and a secondary air stream heated in the cooling water return circuit in the turbine island using an air heater.

As a final concluding point it is worth noting that the overall cost effect of coal treatment can be significant and the efficiency gains/losses in this area can have considerable impacts on the power plant's overall economics. Correspondingly preparing ultra-clean coal, although resulting in an increase in the overall cost of coal preparation and ash handling is under investigation since significant associated cost reduction benefits in the downstream boiler and flue gas cleaning processes are expected [33]. Thus a detailed case specific study on the coal treatment is strongly recommended for any techno-economic study done on coal power generation.

2.2.1.2 Ash handling

There are substantial amounts of non-combustible impurities in coal. As the fuel travels through the high-temperature zone in the furnace, volatile matter and carbon are burnt off whereas the mineral impurities are carried in the form of ash or bottom ash. These ash particles fuse in the combustion zone of the furnace; however when leaving the combustion zone the molten ash is cooled rapidly and is solidified as spherical glassy particles. Ash consists primarily of oxides of silicon, aluminum, iron and calcium. Magnesium potassium, sodium, titanium and sulphur are also present to a lesser degree [41]. Ash formation is a complex process in which the coal type, condition of coal crushers, oxidant, and combustion kinetics play a significant role. This by product ash produced is present in two forms namely: fly ash and bottom ash. Both types are formed as a consequence of the ash melting and cooling/solidifying process.

Fly ash is smaller in size and is carried along with the flue gas into the downstream processes. It is captured through particulate removal processes, which are most commonly either electrostatic precipitator units (ESP) or filter bag houses. Bottom ash is usually larger in particle size and does not fluidize or exit the furnace along with the flue gas; rather it is collected from the bottom of the furnace, and is normally disposed off to landfills. A potential market for fly ash is the cement industry and as for the bottom ash, it is utilized in road way construction as ash mound. It should be noted that the quality of ash is determined by several combustion/coal related parameters such as the loss on ignition, fineness and chemical composition and uniformity of ash which should be taken into consideration if marketing this by product is to be pursued; for further detail on ash formation, composition, quality and possible markets please refer to the CII-ITC Centre of Excellence for Sustainable Development's web site [41].

2.2.2 Air separation unit

The air separation unit has shown to be one of the most expensive components of oxy-fuel power plants. This is due to the large quantities of oxygen required for the oxy-fuel combustion process. A comprehensive report studying oxygen production technologies conducted by the IEA [1] claims that roughly 19.5 tonne/day/MWe oxygen is required for oxy-fuel combustion at concentrations of 95-97% oxygen. In other words for a 1,000 MWe power plant 20,000 tonne/day of oxygen is required. It should be noted that this number would change depending on the percentage excess oxygen deemed appropriate for proper combustion along with the degree of air infiltration allowed into the boiler and downstream flue gas treatment processes.

2.2.2.1 Oxygen production technology alternatives

2.2.2.1.1 Cryogenic air separation

The most proven technology with relatively high capacities for pure oxygen production is the cryogenic air separation technology. According to the IEA report [1], current single train technologies have production capabilities in the range of 4000 tonne/day and new facilities

proposals are in the range of 6000 tonne/day; similar ranges were reported by other sources [14]. Cryogenic air separators' operation can be summarized as follows:

1. Air is filtered to remove all particulates, then cooled and compressed to high pressures;
2. It is then cleaned from components such as CO₂, hydrocarbons, water vapour & nitrous oxides;
3. Air is then compressed and cooled to cryogenic temperatures; and,
4. Two or more columns operating at different pressures and cryogenic temperatures, along with several heat exchangers perform the separation and purification of oxygen.

The fourth step is the actual separation step and the most complex; in this step two or three columns are used in order to separate the cleaned air into high purity oxygen and nitrogen streams (referred to as the “Cold box”). The process is highly integrated in terms of heat recovery as is evident from the presence of high duty heat exchangers such as condenser/re-boiler combinations employed for the columns' operations. There is commonly a higher pressure and temperature column which is fed with air producing nitrogen rich and oxygen rich streams. The nitrogen rich stream after being cooled provides the reflux for the high pressure and temperature column as well as the other column, the low pressure and temperature column. The oxygen enriched stream provides the low pressure column with a feed stream. An intermediate side draw is also present taking some of the air from the high pressure column, sub cooling it and finally directing it as an intermediate feed stream to the low pressure column. Liquid oxygen is then extracted from the bottom of the low pressure column and a nitrogen rich stream is extracted from the top. This nitrogen can be employed in the main air heat exchanger/compressor of the ASU as well as providing further duties for cooling in CO₂ capture and compression process and for the drying of coal as shown by the Canmet Energy study [21]. The liquid oxygen also offers similar duty contributions, since it needs to be vaporized and prepared for boiler conditions, both in terms of pressure and temperature.

2.2.2.2 Oxygen production using ion transport membranes

An alternative technology which has received attention for oxygen production in the recent years is air separation using the so-called ion transport membranes (ITM and OTM). Ion transport membranes are highly selective ceramic membranes. Oxygen from the air ionizes on the surface of these membranes. It then diffuses through these membranes providing a 100% purity oxygen stream. The air needs to be heated to temperatures above 800°C and compressed to approximately 14 bars for this process's proper operation. High recoveries, ≈70%, are possible with this method. Gas heaters are employed to recover the heat from the hot oxygen lean air stream. However, there are still various issues with reliability and durability along with low tested capacities. Also the issue of integration of this process into the overall plant configuration is of great significance. Since a large heat source is required for the heating of the air prior to separation, natural gas furnaces and gas turbines have been proposed by recent studies [1][16]. Thus heat recovery and process integration into the overall scheme is inevitable if these processes are to become competitive. The scaling up of these units might cause serious operational issues that are still unknown. The largest capacity membrane according to the IEA report [1] under testing is at a 0.5 tonne/day limit range and the largest proposed project using this technology would be its application to an IGCC plant to be started up in 2012 with oxygen requirements in the range of 2000 tonne/day. The study also claims that this technology's benefits would be more suitable for the IGCC case and recommends the application of cryogenic air separation for the oxy-fuel combustion plants based on the current state of technologies.

It should be noted that from the operational point of view, the two aforementioned alternative technologies are completely different in nature. In terms of operating temperatures, ITM operates in the 800-900°C range while cryogenic air separation takes place in the -180°C temperature range. Furthermore in terms of energy consumption the ITM requires large quantities of heat input while cryogenic air separation technique consumes significant quantities of high quality shaft/electrical energy. Thus when it comes to the integration and optimization into the overall energy production process a detailed

analysis is required. Comparisons performed by IEA GHG [1] and an earlier MIT study [12] have claimed that 20%+ capital cost reductions are possible with switching to ITM for oxygen production requirements. It should be noted, however, that when considering the energy and fuel requirements of the ITM process, the former study reports comparable costs. A study by NETL [16] compares the cryogenic air separation and the ion transport membrane process applied to a supercritical oxy-fuel combustion process. The study reports that the capital cost of the ITM by itself is approximately 30% less than that of the cryogenic air separation capital cost. However the addition of the expansion turbine and generator used to recover heat and power from the heated nitrogen rich air stream although increases the gross power output it also adds to the overall oxygen production capital costs. The study claims that the largest contributor to the increased costs is the cost of fuel both natural gas and coal employed to provide the heat duty required by this novel, yet immature, technology (ITM/OTM). The auxiliary power consumption of the process for the air preparation using ITM is much higher than for the cryogenic alternatives. Overall the study finds that the overall cost when considering the fuel costs is larger for ITM over its cryogenic oxygen production counterpart. Consequently, in the near term the only practical and cost efficient choice appears to be the cryogenic air separation technology.

2.2.2.3 Optimum oxygen purity

Oxygen is purified so that incondensable gases present in air (mainly nitrogen 79%) are removed prior to the combustion process. These inert gases, however, may leak into the boiler and downstream flue gas treatment processes via air infiltration. The optimal degree of purification required via the air separation unit is determined based on the air infiltration rates, the cost of removal of impurities in the air separation unit and the cost if they are to be removed in the CO₂ purification/compression stages. The cost of purifying oxygen to >99% purities is very high; however this cost drops significantly when lower purities (~95%) are required. In most recent studies there appears to be an inclination towards imposing lower oxygen purity requirements in the front end at a cost of implementing more stringent inert gas removal requirements on the back end.

The earliest study assessing the optimum oxygen purity issue has been a Japanese study [36] which reported the optimal purity to be 97.5% for a supercritical oxy-fuel power plant. The study performed an optimization of the cost of removal of inert gases via the CO₂ compression stages and the ASU's separators in order to find the optimal purity. An IEA study [14] claims the cost of inert removal to be much cheaper downstream along with the CO₂ compression stages rather than in the front end ASU. Other earlier studies have evaluated higher required purities such as 99.5% [40] and 99% [31][44]. However, recent studies appear to have considered an oxygen purity of 95% [3][9][13][14][19][26][37][46]. The logic behind this preferred lower purity is stemmed from the high air infiltration rates in the downstream processes and through the boiler. Several reports [14][19] have considered boiler operation at slightly positive pressures to overcome this issue. However according to both of the aforementioned studies to avoid danger to the operators from leakage of hot gases, dust and carbon monoxide this operating mode is discarded. This mode of operation has only been shown to be possible for small boiler operations [19]. Also another IEA study [14] claims that even for an unusually gas-tight newly commissioned power plant a decrease in gas containment integrity is unavoidable during the major overhauls and also due to thermal distortion and cracking. It is worth noting that the amount of air infiltration in the downstream processes most commonly at the ESP often exceeds that taking place in the boiler.

2.2.2.4 Oxygen distribution

By oxygen distribution we are referring to how much and through which streams the oxygen should be carried and mixed with the combustion gases entering the boiler. The various possibilities commonly considered for oxygen to be introduced into the boiler are:

- Direct injection at the burners;
- Primary recycle stream; and,
- Secondary recycle stream.

Different distribution schemes are proposed by different sources. The only common consensus appears to be that using the primary recycle stream is not recommended this is due to the possibility of fires and explosions in the mills [14].

Introduction of oxygen into the secondary recycle stream however also has limitations. This is due to ducting codes related to the oxygen content of streams and if the same ducts used in air firing of coal are to be used for the case of oxy-fuel these codes must not be violated. This limit is claimed to be in the range of 23-40% by volume concentrations of oxygen [14] if the ducts used in air-firing are to be utilized. Streams having oxygen concentrations greater than 40% need to be treated as pure oxygen streams for safety reasons. Consequently it can be seen that some of the oxygen is bound to be introduced at the burners. The same study notes that direct injection of oxygen could be a powerful tool for NO_x emission control as well, however, further research was recommended.

2.2.3 Boiler, steam cycles and flue gas recycle

2.2.3.1 Boiler/steam cycle types

In pulverized coal fired power plants the efficiency of the power plant is increased when it is operated at higher steam temperatures and pressures, which can simply be explained by the fact that a heat engine's theoretical efficiency is a direct function of the temperatures of the heat engine's hot and cold bodies. Based on the steam cycle operating conditions boilers are categorized in the following approximate manner:

- Sub Critical: Operated at 12.4-16.5 bar and low 500°C 's with ~35% net electrical efficiencies [4];
- Super Critical (SC): operated at low 200's bar and mid 500°C 's with ~40-45% net electrical efficiencies; and,
- Ultra Super Critical (USC): operated at low 300's bar and 600°C's with ~49% net electrical efficiencies [6].

However the above categorizations can differ slightly from study to study, an example of this deviation is observable in the categorization done by EPRI in Figure 1. This figure exemplifies the ranges of operating temperatures for the different stream cycles mentioned. Another categorization of pulverized fuel power plants is often employed referring to an “advanced supercritical” or “advanced ultra super critical” as shown in Figure 1. According to an NETL study [16] the word advanced refers to next generation conditions chosen to be in agreement with industrial consortiums for advanced material development. The operating conditions for these power plants typically fall in the range of pressures employed for ultra supercritical plants along with temperatures in the range of mid 700°C and possibly higher temperatures. Sub critical and super critical power plants are in operation along with few ultra supercritical plants. According to the world coal institute, higher efficiencies, up to 50%, are attainable with the ultra super critical technology. Denmark, Germany and Japan have been focusing on this technology and large amounts of research is being conducted in the area of developing corrosion resistant high alloy steels for these purposes [45].

Thus most recent studies tend to focus on the simulation and study of oxy-fuel combustion at super, ultra super and advanced super critical conditions, due to their higher attainable efficiencies. It has been shown that retrofitting subcritical power plants is not economically feasible due to their inherent lower efficiencies. A study [39] addresses this issue comparing the economics of advanced super critical and sub critical power plants firing bituminous coal. Thus the tendency in most recent literature has been to analyze higher efficiency super and ultra super critical cases [24][33].

It should be noted that there is an advantage for the case of greenfield power plants if oxy-fuel is used. This is due to the possibility of boiler size reduction. For the case of greenfield plants, in order to reduce the recycle rates, a slimmer and longer convection section would be required to maintain suitable velocities as proposed by an IEA study [14]. In other words the same amount of heat is produced but much less combustion flue gas is present to absorb it all, which results in much higher temperatures.

There are a number of factors that affect the economics of oxy-fuel power plants. Commonly, oxy-fuel power plants' performance and economics are evaluated by comparing them to conventional air fired coal power plants as an appropriate without capture benchmark. Generally the cooling water requirements of the oxy-fuel power plant is 1.2-1.5 times those of the air-fired power plant and this factor can change depending on the degree of energy integration minimizing the increase in water requirements of the power plant [46]. Furthermore, studies have done comparisons between capture ready and with capture cases for which the issue of the definition of capture ready power plant is another important factor. For instance the inclusion of the boiler modifications in the definition of capture ready power plants would result in a 5-10% reduction in the levelized cost of electricity (LCOE) as reported by a study done by the formerly known DTI [46].

2.2.3.2 Flue gas recycle configurations

The different configurations proposed in the literature are commonly categorized either as wet or dry or in some cases a compromise of the two. The distinction between the dry recycle and wet recycle is based on the location of the recycle stream relative to the main condensing heat exchanger. However this does not appear to be an absolute distinction since there appears to be a required compromise between the possible energy efficiency gained from recycling the flue gas in the wettest conditions and the acid gas corrosion issues, ESP and milling unit operation difficulties which can be avoided by using dryer, cooler and cleaner recycle streams. A factor that is important in this process selection is the coal composition since the combustion flue gas composition is directly dependent on it.

In summary several factors that are of great importance in choosing the most suitable configuration are:

- Energy savings: level of energy savings from reduced cleanup and cooling & reheating processes;

- Operability issues: mill and ESP operation difficulties associated with each configuration;
- Equipment and construction material's limits: level of acid gas corrosion expected depending on coal type;
- Safety concerns: hot particulate removal & coal milling operations and fires caused by residue build up from dirty recycle;
- Oxygen distribution: locations where the mixing/introduction of the oxygen can be done efficiently and safely; and,
- Operational flexibility: capability of the overall process and its response when possible perturbations occur in operational parameters.

An IEA study [14] performed on a power plant burning black coal considers several configurations, entailing various degrees of cooling and reheating of the primary and secondary streams. The authors mention that in theory there appears to be various possible recycle configurations. However, when the constraint for the coal type and the corresponding flue gas composition is considered, there is actually little flexibility in terms of how the primary stream is treated. According to the author the primary stream must be cooled and scrubbed to remove moisture and soluble acid gas components such as SO_x and HCl and then be reheated (250-300°C) before being fed to the mills. This is to avoid possible damage to the recycle fan and the complex mill operation as well. The secondary stream, however, can be recycled without drying at higher temperatures. The most reasonable secondary recycle conditions between all the choices was concluded to be cooling the secondary recycle to temperatures above the acid dew point (~160°C) with no moisture removal. It is then recycled through the gas/gas heater. The ESP temperature is in the 230-270°C range and is determined by heat balance which depends mainly on recycle flow and the economizer gas exit temperatures. The study reports that this design is the optimal compromise minimizing the quantity of gas cooled and reheated, guaranteeing low moisture content going to the milling plant with maximum utilization of the conventional plant equipment. Also it is noted that this concept is applicable to normal start-up with air

firing. Conversely the mentioned disadvantages were the performance and operability issues of the elevated ESP temperatures and the higher impurities present in the boiler flue gas passes. Furthermore a maximum temperature limit of 250-300°C for the primary recycle stream needs to be in place to ensure proper operation of the mill bearings. The quantities of acid gases present in the boiler are a big concern since they are almost five-fold of those reported for conventional air-fired cases. It is recommended that more experimental data is needed since the accelerated metal wastage of side wall tubes and the gas side convection section of the boiler is not ignorable. Also a careful consideration of the surface temperatures of metals is mentioned to be a must. Acid corrosion concerns are also expressed for the complex operation of the mill's machinery. Other corrosion problems mentioned by the authors are those related to the main flue gas heater at the inlet of the primary recycle stream where temperatures below the acid dew point is reached. The same concern is present for the primary recycle fan, ductwork and water removal system as well.

Alternatively another IEA study [19] reports a recycle configuration that involves a single stream coming out of the electrostatic precipitator. After being cooled through the main gas/gas heater a direct contact scrubber is employed to cool the 80°C boiler gases, the scrubber is equipped with a dryer at the top. The recycle stream is separated at this point by the recirculation fans. One of the streams is then introduced to the coal pulveriser where it is mixed with coal and entrained into the boiler. This way, some of the acid gases are condensed in the direct contact scrubber. However, the concern for the corrosion issues in the main gas/gas heater is not addressed.

Another study by the Department for Business Enterprise & Regulatory Reform (BERR) assessing CCS technologies for the Canadian market proposes a dual recycle process for lignite-fired ASC oxy-fuel power plants [46]. The recycle streams consist of a primary and a secondary recycle stream. The secondary recycle stream is extracted after flue gases are cooled in the main gas/gas heat exchanger, cleaned from particulates and further cooled in a heat recovery unit. In addition, the primary recycle stream is further cooled in another heat

recovery unit and its sulphur oxides content is reduced through an FGD unit. It is then further cooled via a direct contact cooler prior to being sent to the mills for fluidization and carriage of the fuel into the boiler. The authors claim with this arrangement that the danger of high temperature gas side corrosion for oxy-fuel firing is comparable to that of air firing. Also sufficient reductions in particulate and acid content of the primary flue gas recycle stream are achieved for proper operation of the mills and avoidance of corrosion problems [46].

In summary, there appears to be the realization that presently the most reliable configuration is to at best recycle a major portion of the flue gas in wetter/hotter conditions in order to minimize heat loss, while cleaning the remaining portion of the flue gas from acid gases and reduce its temperature to acceptable levels for drying, fluidizing and carrying the fuel to the boiler. Future developments and improvements in material limits, coal milling and hot ESP's operability might assist in the practicality and the safe use of the wetter and hotter recycle alternatives.

2.2.3.3 Amount of flue gas recycle required

Various boiler modeling and simulation studies indicate that if roughly 2/3 of the flue gases by volume is recycled, similar combustion and furnace outlet temperatures to the air fired cases are achievable. The furnace outlet temperature limit is normally specified by boiler manufacturers and is based on the ash softening temperature. However, when an insufficient amount of recycle gas is provided this temperature limit is exceeded causing slagging and ash deposition problems to occur.

A study by NETL [16] determines the required recycle stream amounts to be 70-72% of the flue gases exiting the boiler. Most recent studies propose recycle rates in the range 65-72% of the flue gases [14][16][19][46].

2.2.4 Flue gas clean up

Flue gas clean up processes for a conventional air-fired pulverized fuel boiler are generally comprised of ESP or bag house, dehydration/condensation processes, deNO_x and deSO_x units. However, oxy-fuel combustion processes require further dehydration and inerts removal processes. Most studies in the late 1990's and early 2000's included the gas clean up units for oxy-combustion as required for the case of air combustion specially deSO_x units but with a reduction in size. The reasoning was mainly the corrosion problems related to the presence of the aforementioned acid gases in the flue gas. Some studies suggest that there is an optimal percentage volume of the recycle gas that needs to be desulphurised and the rest can be recycled as is [31]. However, there is a possibility of eliminating the acid gas removal processes in the case of oxy-fuel combustion. The issue of acid gases can be dealt with, if the temperature of the recycle streams is maintained above the acid gas dew point. For instance, if ESP's could be operated hot and an indirect method of coal drying was in place as shown previously using the nitrogen from the air separation unit, and minimal cooling and reheating was present, the NO_x and SO_x removal units could be eliminated with small anticipated corrosion issues. Another issue that might impose acid gas removal unit requirements is what is intended to be done with the captured CO₂ and what purities are expected to be achieved for this stream. For enhanced oil recovery purposes any NO_x present would impose undesirable results due to the process's low NO_x tolerance limits. The aforementioned issue ties the flue gas cleanup process configuration closely with the CO₂ cooling and compression process schemes selected.

An IEA study [14] considers both cases and proposes that neither an FGD nor a NO_x removal process is required. Instead a simple cryogenic separator and, depending on what purity of CO₂ is required, a two or three stage flash separators or one with a distillation stage is sufficient. The study found that producing the CO₂ purities of 98% vs. the base case of 95% does not significantly affect the costs however there will be a 2.3% reduction in the amount of CO₂ captured (the recovery decreases to 88.4% from 90.7%). It is believed that performing the purification of the flue gas is much cheaper if done at the back end rather

than investing on the production of higher purity oxygen in the front end. To achieve higher purities (98%) another flash separator along with a compressor was added (the compressor is set in place to reduce capture loss). Capital costs are slightly increased due to the extra compressor, separator and the piping. Much of the SO_x and NO_x are re-circulated back to the boiler and the rest are captured with the CO_2 . It is simple and cheap to extract these components during the CO_2 purification which already has to use cryogenic flashes to separate incondensable components such as O_2 , Ar and N_2 , and the author claims that an extra distillation step is all that has to be added if the option of storing them with CO_2 is not possible. A study by NETL [16] quantifies these savings to be negligible for the case of increasing the required oxygen purity limit from 95% to 99%. The study recommends performing the purification further on downstream with CO_2 compression processes with increased auxiliary power requirements as well. The study claims that the offset is 9% of the ASU capital cost for the increased purity requirements.

Similar results are reported in a recent study by John Davison [13] where a cryogenic separator was proposed and it was mentioned that some of the acid gases will also be removed in the cryogenic separator and the remaining will be co-captured in the CO_2 stream. The author mentions that the storage of CO_2 containing SO_x and NO_x has not yet been demonstrated, however, recent work shows that 90% of these gases can be converted to nitric and sulphuric acid and can be easily removed, but implications to the plant design and materials are expected.

Another IEA study [19] considers three options:

1. Scrubbing with amine which was found to be costly as expected;
2. Compression of the gas at high pressure: 90% CO_2 purity was achieved but too high of O_2 content was present for pipeline transportation; and
3. Chilling of the gas, compression and cryogenic separation: through which the required purities were achievable.

The third option which was chosen involves direct water scrubbing to cool the flue gases and remove the remaining particulates downstream of the ESP. The scrubber is equipped with a dryer at the top. The recycle stream is then separated at this point by the recirculation fans and is heated prior to entering the boiler.

A comprehensive study by NETL [16] considers various scenarios dealing differently with flue gas clean up. From their study it is concluded that the presence of the FGD unit is a must for most black coal types due to their usually higher sulphur contents. The study was performed on an Illinois #6 coal, and it was determined that the presence of a wet limestone FGD unit is inevitable. The study determined that due to the 60-70% reduction in NO_x no SCR was required. Bag filters appeared to be the choice of particulate removal technology. Co-sequestration in the main cases considered was assumed to be feasible as well.

A study by BERR [45] reaffirms the fact that the requirement of an FGD unit is solely dependent on the coal type. There are two factors of importance in regards to the coal composition, one is the sulphur content and the other is the heat content. The reasoning behind heat content affecting FGD requirements is due to the fact that for the case of lignite in order to provide the same thermal output as the sub-bituminous cases a fuel input rate of 2.5 times those of the sub bituminous coal cases is required.

2.2.4.1.1.1 Acid gases

An IEA study previously introduced [14] claims that although the level of acid gas condensations seems uncontrolled for the case of oxy-fuel combustion (due to the absence of dedicated NO_x and SO_x removal processes) they are in fact somewhat well managed. This claim is due to the fact that there is no tendency for any acid gases or water to condense in the coal milling operations since there is sufficient dehydration present to avoid acid gas corrosion. It is also adequately noted that NO_x emissions are inherently much smaller for the case of oxy-fuel ~50% of the air-fired case. Furthermore 70% of the SO_x produced is believed to be returned to the boiler through the primary and secondary recycle streams and another 5% is believed to be scrubbed via the direct contact cooler columns. The final amounts of NO_x and SO_x contributions to the compressed CO₂ stream are calculated to be 26 mg/MJ and 534 mg/MJ respectively, which are not expected to impose any significant hardware corrosion problem.

Another study by IEA [19] claims that limits of NO_x can be met with proper staged combustion and maintaining low temperature at furnace exit, thus no SCR is proposed and the remaining NO_x will be condensed with the CO₂ stream. No FGD is proposed since the recycle stream is reheated and the acid dew point corrosion problems can be avoided and the remaining SO_x in the captured stream are going to be condensed as well along with CO₂. This study, however, expresses concerns about acid gas corrosion in the gas/gas heater and advises that corrosion resistant material be employed. The issue of recycle fan, and boiler corrosion problems however are not addressed.

Conversely the study by NETL [16] claims that, considering a 72% recycle rate, the amount of SO_x present in the flue gas increases by a factor of 3.5. Thus a coal with 2.5% sulphur content would result in 8.75% sulphur concentrations in the boiler. This exceeds the boiler material design limits of 3.5% sulphur content to avoid excessive corrosion. The study proposes that if the coal sulphur content were at most 1% then the elimination of the FGD unit is a possibility, otherwise its elimination is not advisable. The study proposes that the

cost savings achieved by eliminating the FGD unit in terms of a percentage of the cost of electricity is in the range of 8.5%.

2.2.4.1.1.2 Particulates

In most recent literature electrostatic precipitators have been the choice of particulate removal technology having very high efficiencies. Hot ESP's are in most studies avoided due to operability difficulties. However due to the higher efficiencies achievable by minimizing cooling/reheating processes in the recent studies, they are gaining popularity. In general particulate emissions from oxy-combustion process are further decreased due to presence of the recycle stream and the downstream clean up processes. For most studies these emission are close to 0 mg/MJ [14].

A study done by University of Athens [26] proposes the use of a combination of two hot and one cold ESP units. In this study the recycle stream is first treated in a hot ESP, partially mixed with oxygen then heated with the remaining flue gas which is further treated in a cold ESP. The authors explain that the necessity for a second hot ESP is due to the absence of a gas/gas heat exchanger in the simulated process [27].

In contrast, the study by NETL [16] considers fabric filters as the choice for particulate removal. The study proposes that higher particulate collection efficiencies are possible for the case of oxy-combustion compared to the case of air-fired combustion. The reasoning behind the previous statement is the increased density of the flue gas and lower flow which causes longer residence times and thus higher collection efficiencies for fabric filter when applied to oxy-fuel combustion.

2.2.4.1.1.3 Mercury removal

Mercury is a potential corrosion enhancer in the CO₂ purification sections. An IEA study [14] proposes that an adsorption process can be put in place after the 30 bar compression stage using charcoal impregnated with sulphur to fulfill the mercury removal task.

Another IEA study [19] reports that EPRI has proposed the best technology to be adsorption using activated carbon injection and then removal via bag filters. The best configuration proposed is for this process to be placed after the electrostatic precipitator where larger particulates have already been removed and bag filters can be employed for the removal of smaller size activated carbon particles and other particulates.

The study by NETL [16] mentions that for the cases of ESP and FGD employment the higher oxidized and elemental mercury content of the flue gas results in higher removal efficiencies. This study proposes that 90% removal is performed via various downstream processes: A portion of the mercury is absorbed by the unburnt coal particles and is collected in the bag house. Another portion of the mercury is collected in the wet scrubber and with the addition of a cheap additive to these units the mercury removal efficiency is enhanced resulting in an overall combined 90% mercury removal.

2.2.5 Flue gas compression

Flue gas compression is claimed to be one of the most energy intensive processes in oxy-fuel combustion due to its high compression and refrigeration requirements.

Various cycles have been proposed in the recent literature [14][16][19][40][47]. The CO₂ compression process in most recent studies is integrated with the inert gas removal, and the optimum conditions desired for this process affect the costs significantly. Requirement for transport and further compression to high pressures such as the removal of any moisture also affect the costs. The desirability of achieving higher purities of CO₂ at the cost of CO₂ recovery reduction is another important factor in the design of these cleaning and compression units. The higher the pressure, the purity and recovery combinations required, the larger the imposed refrigeration and compression duties will be. The type of compressors proposed by most studies appears to be adiabatic. This process selection is due to the possibility of heat integration/recovery via inter-stage coolers with streams that require heating such as the boiler feed water or nitrogen air streams used for coal drying.

The process scheme selection and consequently the costing of these processes are dependent on the following considerations:

- Purity limit proposed;
- Recovery limit required;
- Temperature and pressure requirements for transport and storage purposes; and
- Heat integration schemes proposed.

Detailed simulation and design of process components is a must. This is due to commonly optimistic assumptions made for the cooling medium temperatures, which is dependent on whether a cooling tower or an open loop using sea water is going to be employed to provide the cooling. Also integration of these units to recover the low grade thermal energy available at the adiabatic heat exchangers for heating upstream streams of the compression unit requires thorough assessments. The issue of thermodynamic limitations, fluctuations and perturbations in flow and operating conditions and their effects on the performance of such arrangements appears to not have been addressed in a convincing level of detail.

An IEA study [19] proposes the following cycle to prepare the flue gas after being scrubbed in a direct contact cooler for storage. In the CO₂ separation section the gases are compressed in multiple stages with inter-stage coolers employing boiler feed water and also IP/LP condensate as the cooling medium. The gases are then compressed to 30 bars. After compression the flue gas is dried in a desiccant drier to reach a low dew point (-60°C) to prevent ice formation. It is then forwarded to the cold box where inert removal using two flash separators is performed operating at -26 and -55°C, respectively. Then, the CO₂ is evaporated to provide some refrigeration and subsequently compressed to 110 bars and ambient temperature. The inerts leaving, which are at high pressure, are passed through an expander to recover some electrical energy. The authors also mention that heat exchangers in this scheme are required to be constructed from corrosion resistant materials in order to avoid acid corrosion issues. In terms of energy consumptions the power consumption of their unit is approximately 10.8% of the power plant's gross electric power output with 93% CO₂ recovery capabilities.

Another IEA study [14] proposes similar cooling and compression stages. Initially by treating the flue gas via a combination of direct and indirect contact see water coolers. The flue gas is then further compressed and cooled in multiple stages until it reaches a pressure of 30 bars after which point it is dried and is further cooled and compressed through multiple cooling and expansion stages to provide the cooling duties required, followed by further compression of the flue gas to 110 Bars and 40°C. It should be noted that once CO₂ reaches a liquefied state its pressure can be increased simply by pumping it, which relatively consumes very little energy. The capture and compression in this study is proposed by Air products and the power consumption reported is 8.8% of the power plant's gross power output.

A third study done by the DOE [16] for which Air Liquide has proposed the capture and compression design involves multi-stage compression and cooling along with drying, and the electrical energy consumption of the design is reported to be 8.35% of the power plant's gross power output.

2.3 Degree of integration

One of the most crucial tasks to increase the efficiency of oxy-fuel combustion processes is to make the best use of the large quantities of low temperature heat available. This heat is available in the air and flue gas which need to be cooled in the adiabatic compressors' inter cooling stages of the ASU and those of CO₂ compression units respectively. The heat sinks that can make use of the aforementioned heat sources are the flue gas recycle streams that need to be reheated prior to going back to the boiler, the boiler feed water and/or the nitrogen stream used for indirect drying of coal in some of the previously discussed configuration schemes. Normally, steam is bled from steam turbines to heat these streams; thus there exists an opportunity for some efficiency gains by replacing this bled steam via heat recovery processes. In the IEA study [14] for instance the aforementioned required heat is partially attained from the combined primary recycle and the product gas streams, along with much larger quantities of heat that are recovered from the after coolers of the ASU and CO₂ adiabatic compressors. The heat recovered from these sources is subsequently utilized in downstream endothermic operations. In the aforementioned study 172 MW_{th} of heat is recovered in total. The breakdown is as follows: 55.3 MW_{th} from the ASU, 66 MW_{th} from the CO₂ compressors and 51 MW_{th} from the flue gases. This study also mentions another source of integration to be direct mechanical drives, i.e. taking shaft power from the turbines. However, very few studies have even considered this option since it is a heavily involved and complex one. Furthermore due to the highly complex steam cycle and the large number of unknowns for the oxy-fuel start up process, it is not considered to be a practical option with the current number of unknowns and the relatively immature stage of the technology. Table 2 provides a more detailed breakdown of the integrated system presented by the aforementioned study [14].

Table 2: Summary of heat integration processes [14]

Heat Source	Amount (MW _{th})	Heat transferred to
ASU adiabatic compressors	55	Condensate
CO ₂ adiabatic compression (early and later stages)	50	Condensate
CO ₂ adiabatic compression (early stages)	16	Boiler Feed Water (BFW)
Flue gas	28	Boiler Feed Water
Flue gas	18	Condensate
Flue gas	5	Vented inert gases

In another IEA study [19] nitrogen is used for indirect coal drying since it has a pressure high enough for fluidizing the coal thus reducing the electrical power duty required for an independent air blower. Nitrogen is also heated with the hot water stream from ASU compressors and the CO₂ treatment plant thus recovering some more heat. A breakdown of this heat integration scheme is provided in Table 3:

Table 3: Split heating sources required for the drying of coal [19]

Heat source	Duty (MWe)
ASU	46
Flue gas treatment	96

In the same study the waste heat from CO₂ compression intercoolers are used to heat the boiler feed water and the condensate at the HP/LP feed water heaters, consequently taking away from the duty of IP/LP feed water heaters.

A similar approach for drying is proposed by a Canmet ENERGY study [21] employing cold nitrogen produced by the ASU unit to initially provide cooling duties for the intercoolers of the compression unit and the hot dry nitrogen stream obtained is afterwards used for the coal drying operation.

Another study considering the use of ITM for oxygen production is the only one of its kind quantifying the degree of integration possible [16]. The study indicates that the heated and

pressurized air used as the source of oxygen in the separation process is initially employed for the heating of the incoming air stream and finally is fed to a parallel expansion turbine producing 198MW of power. However to enable this heat recovery the turbine/generator needs to be purchased and added to the process scheme.

In conclusion, there appears to be large areas of possible heat integration from oxy-fuel power plants' oxygen production, CO₂ clean up and CO₂ compression processes in order to achieve larger quantities of power from steam turbines by reducing/eliminating bleed streams conventionally required to provide heating duties. Also in the future possibilities may exist for integration of the shaft power consuming/producing processes.

3 COST COMPARISON OF CARBON CAPTURE TECHNOLOGIES

Performing cost comparison of power production technologies with CO₂ capture capabilities is a task that necessitates detailed simulations of each technology along with similar site specifications, a common financial basis and coal type under study. This requires large amount of resources and might not be a feasible or a reasonable task to complete, in particular when the core deliverable of a project is the assessment of only one technology. Similarly, in this study the evaluation of Advanced Super Critical Oxy-fuel combustion is the main concern, but cost comparisons with other studies and technologies are also desired. In order to fulfill this comparison task and avoid any erroneous conclusions several issues need to be addressed to develop a proper comparison framework. The issues are as follows:

- Financial Assumptions:
 - Interest rates forecasted;
 - Basis year in which the cost developments are performed;
 - Fuel price forecasts; and,
 - Investment schedule.
- Operation assumptions:
 - Length of life of the plant; and,
 - Capacity factor.
- Site specifications:
 - Distance from a large body of water;
 - Distance from the coal mine; and,
 - Distance from the captured carbon storage site (if considered).
- Coal type and choice of technology for the comprising processes/unit operations.

Since in different studies the level of detail and transparency in inclusion and reporting of the above factors differs, to a large degree the task of adjusting these parameters for obtaining a common comparison basis might be very cumbersome, if not impossible. In addition some conditions cannot be adjusted for, such as the effect of the difference in costs due to un/availability of a large body of water in close proximity of the operation or the

effect of coal type. Thus it is unadvisable to perform cost comparisons of results reported by different sources having very different fundamental assumptions. The logical approach would be to look at studies in which the performance and economics of multiple technologies are assessed based on a common financial and assessment methodology basis. In addition, in some literature certain operational assumptions are made, in particular for technologies that are immature, that are not commonly observed in other similar studies, which might affect the overall economics of a technology and needs to be pointed out and considered when drawing conclusions from the results presented.

3.1 Coal power generation technologies

Various technologies are under investigation for power generation with CO₂ capture capabilities by both industry and academia; however three of the more promising technologies are discussed in this report namely: conventional coal combustion with CO₂ capture using amine absorption, integrated gasification combined cycle process with CO₂ capture and oxy-fuel combustion with CO₂ capture.

3.1.1 Conventional Coal Combustion with CO₂ capture

The most proven and mature technology for CO₂ capture from coal fired power plants to date is the post combustion capture process via amine absorption processes. In this technology coal is combusted using the conventional combustion techniques with air. The flue gas exiting the boiler is cleaned from its particulates and sulphur containing compounds. However, instead of being released to the atmosphere via a stack, it is carried to the absorption process. In the absorption process the CO₂ present in the flue gas, which is approximately 10-15% of the flue gas, is absorbed, commonly via a monoethanol amine (MEA) solvent and the remaining inert gases, which is mostly nitrogen gas, is released to the atmosphere. The absorbed CO₂ stream is then compressed and cooled to achieve conditions suitable for either storage or Enhanced Oil Recovery (EOR).

The main drawback of this technology is the large de-rates imposed on the overall power plant output due to large quantities of low pressure steam requirements for regenerating the solute rich solvent, which is thought to be extracted at the IP/LP turbines' cross over pipe. The absorption and regeneration process is a highly optimized one and both industry and academia are working towards inventing new solvents which have high absorption capacities, yet low energy requirements for regeneration, along with low levels of degradation and solvent loss. One of the other issues commonly noted concerning this capture technology is the large footprint requirements for the absorption/desorption process. It is however considered as one of the main contenders for CO₂ capture processes especially for retrofit purposes. A process flow diagram of the overall process is provided below, see Figure 4.

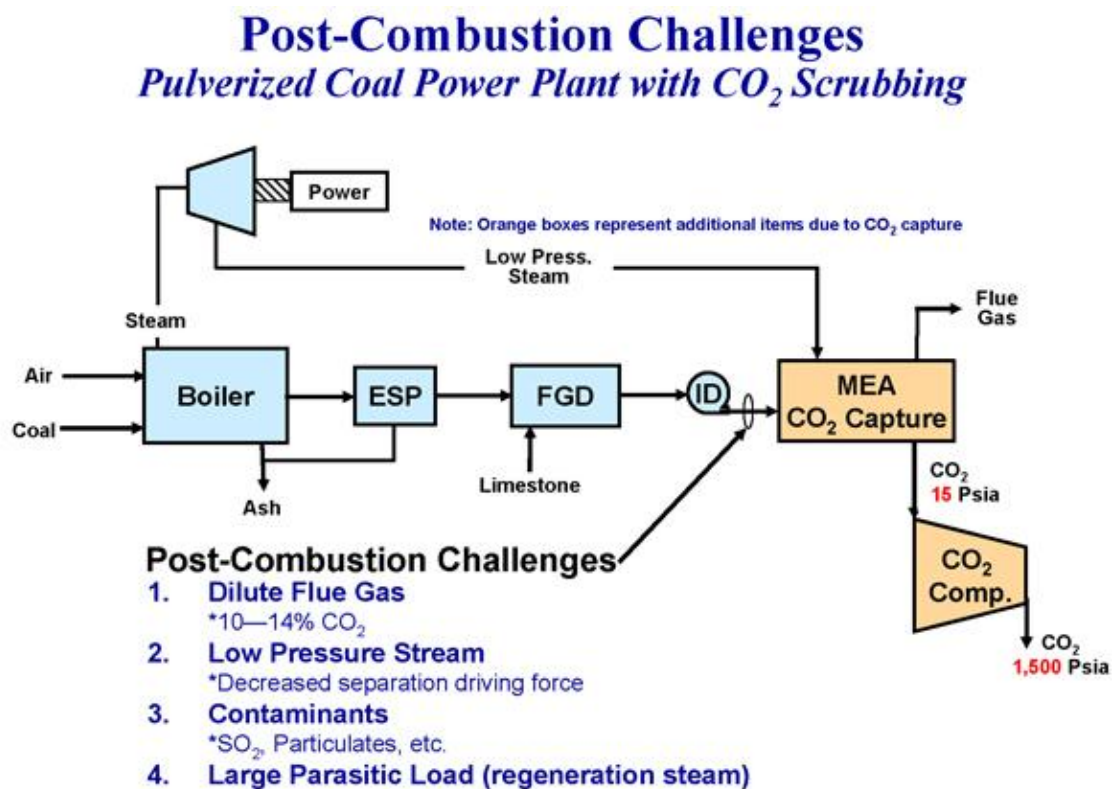


Figure 4: MEA Absorption [35]

3.1.2 Gasification process

One of the coal power generation technologies pursued by several utility companies, which can be integrated with carbon capture units, is Integrated Gasification Combined Cycles (IGCC). In the gasification process a gasifier, instead of a boiler, is employed to convert coal or a hydrocarbon fuel into a synthetic gas, which mainly contains carbon monoxide, hydrogen, hydrogen sulphide and small amounts of methane. A gasifier differs from a combustor in that the amount of oxygen available is at sub-stoichiometric levels so that only a relatively small portion of the fuel burns completely. In this "partial oxidation" process most of the carbon-containing feedstock is chemically broken apart by the gasifier's heat and pressure to produce the syngas [34]. Similar to Pulverized Coal combustion the composition of syngas varies depending upon the operating conditions of the gasifier and the fuel composition. Figure 5, extracted from an NETL study [15] illustrates the difference in flue gas compositions produced as a function of the tonne of oxygen present per tonne of moisture ash free (MAF) coal fed for the gasification and combustion process.

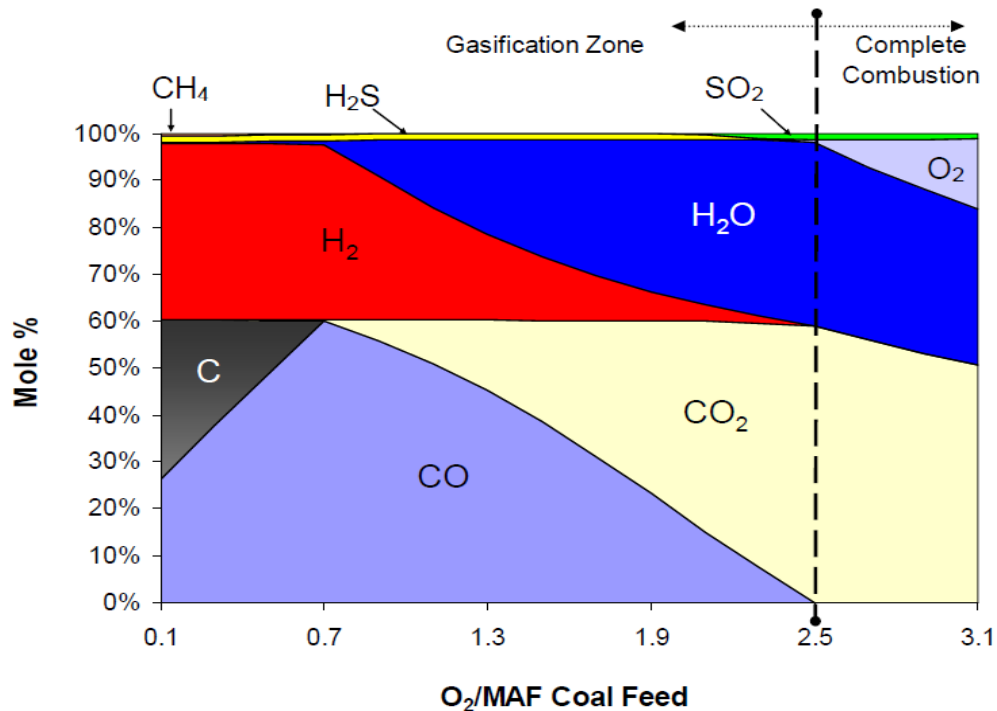


Figure 5: Gasification phase diagram [15]

The syngas exiting the gasifier is commonly referred to as “raw syngas” which needs to be further treated to remove its sulphur containing compounds and convert most of carbon monoxide into carbon dioxide for capture purposes, since CO is not captured well via amine or physical solvent systems. Similar to combustion technologies purified oxygen is used in the gasification process instead of air as the source of oxygen, since the nitrogen present in air dilutes the stream and increases the size and cost of the gasifier and the downstream processes required for CO₂ capture purposes. In IGCC with CO₂ capture processes the syngas is commonly passed through a shift reactor, where carbon monoxide reacts with water vapour to produce H₂ and CO₂. It should be also noted that most of the sulphur is converted to H₂S in the gasifier along with small quantities of COS produced which is converted to H₂S in the shift reactor. Furthermore the reaction taking place in the shift reactor is an exothermic one, thus the heat liberated can be transferred to steam to produce power in steam turbines. The shifted syngas stream is then typically cooled and its H₂S content is removed via a Selexol absorption process, the H₂S free stream, then goes through another Selexol absorption stage where a purified CO₂ stream is extracted for storage. The remaining H₂ exiting the CO₂ stripper is sent to a gas turbine, followed by a heat recovery steam generator (HRSG) to produce steam for a steam turbine to further extract power. The use of these two types of turbines namely a combustion turbine and a steam turbine in combination, known as a "combined cycle," is one reason to why gasification-based power systems can achieve high power generation efficiencies. Currently, commercially available gasification-based systems can operate at around 40% efficiencies (HHV-based) [34].

The H₂S acid gas stream is then sent to a Claus process where elemental sulphur can be produced. As for the particulates in the gasifier, it should be noted that their amount is reduced significantly compared to coal combustion processes, since most of the mineral components in the fuel that do not gasify, such as sand rock and ash, are collected as inert glass-like slag from the bottom of the gasifier. Thus the particulate removal procedures required to treat flue gases from conventional coal combustion processes are eliminated or significantly downsized in the gasification processes. A small fraction of the mineral matter

is blown out of the gasifier as fly ash and requires simple removal procedures downstream [34]. An overall process flow sheet for an IGCC process with CO₂ capture is shown in Figure 6.

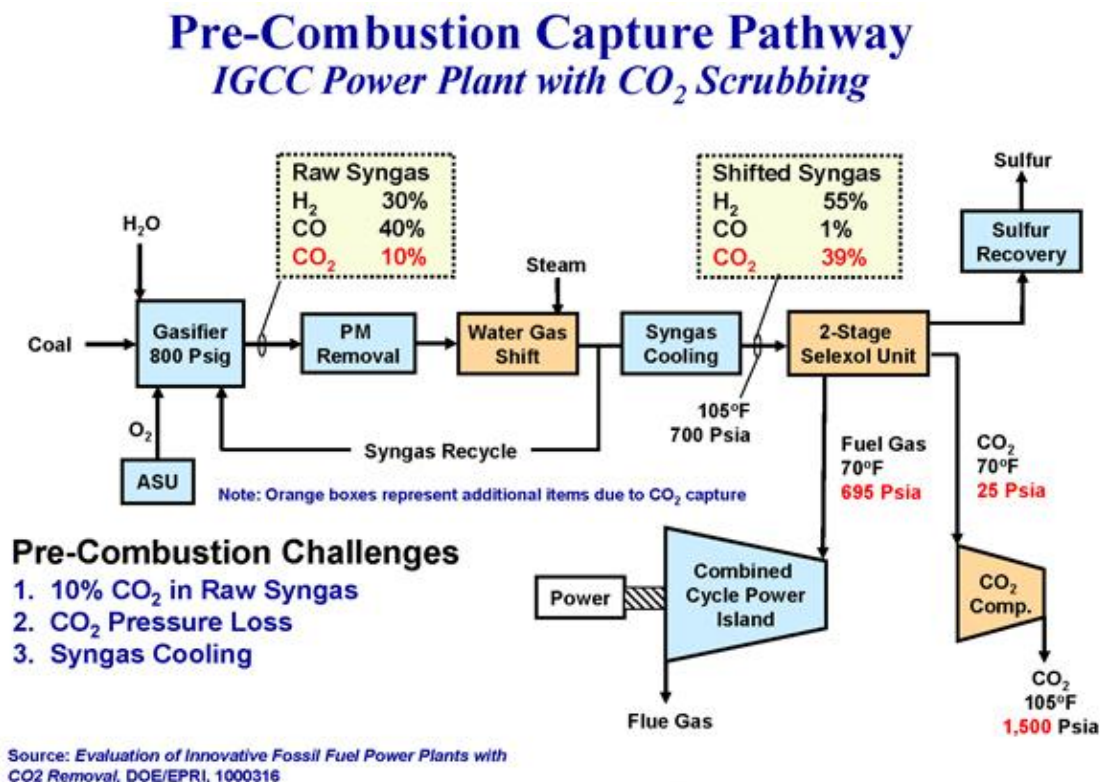


Figure 6: IGCC process with CO₂ scrubbing [35]

Nitrogen oxides are generally not formed in the reducing oxygen deficient environment of the gasifier. There are various types of gasifiers available in the market with significant variations in design proposed by several technology providers. Generally, there are two main types of gasifiers in terms of the fuel feed conditions, dry-feed systems and slurry fed systems. The choice of gasifier is very much dependent on the coal properties. Low density high moisture brown coals are generally not suited for slurry fed gasifiers since the resulting solids contents are too low for proper gasifier operation. Furthermore, generally high ash coals are more suited for dry-feed systems. Table 4 below shows three of the common gasifier types and their operation characteristics and capabilities.

Table 4: Common gasifier types and their corresponding characteristics [8]

Technology Name / Design Feature	GE Energy	E-Gas (Conoco Phillips)	Shell
Feed System	Coal in Water Slurry	Coal in Water Slurry	Dry Coal, Lock Hopper & Pneumatic Conveying
Gasifier Configuration	Single Stage Downflow	Two Stage Upflow	Single Stage Upflow
Gasifier Wall	Refractory	Refractory	Membrane Wall
Pressure (psig)	500-1000	Up to 600	Up to 600
Slag removal	Lock Hoppers	Continuous Pressure Let Down	Lock Hoppers
Syngas Cooler Design	Downflow Radiant Water Tube / Horizontal Convection Firetube	Downflow Firetube	Downflow Water Tube Design
Particulate removal	Water Scrubbing	Hot gas filter at 350-400 C	Hot gas filter at ~250 C
Recycle Gas to Gasifier	None	Clause Plant Tail Gas to Gasifier	Recycle to Quench Raw Gas to 900 C (Radiant HR)
Fine Recycle	Scrubber Fines to Slurry Preparation	Filter Catch to 1st Stage Gasifier	No Fines Recycle. Filter Catch Sold

Figure 7 is a histogram of some of the common gasifier types illustrating each technology's popularity considering technologies in operation and also future planned projects.

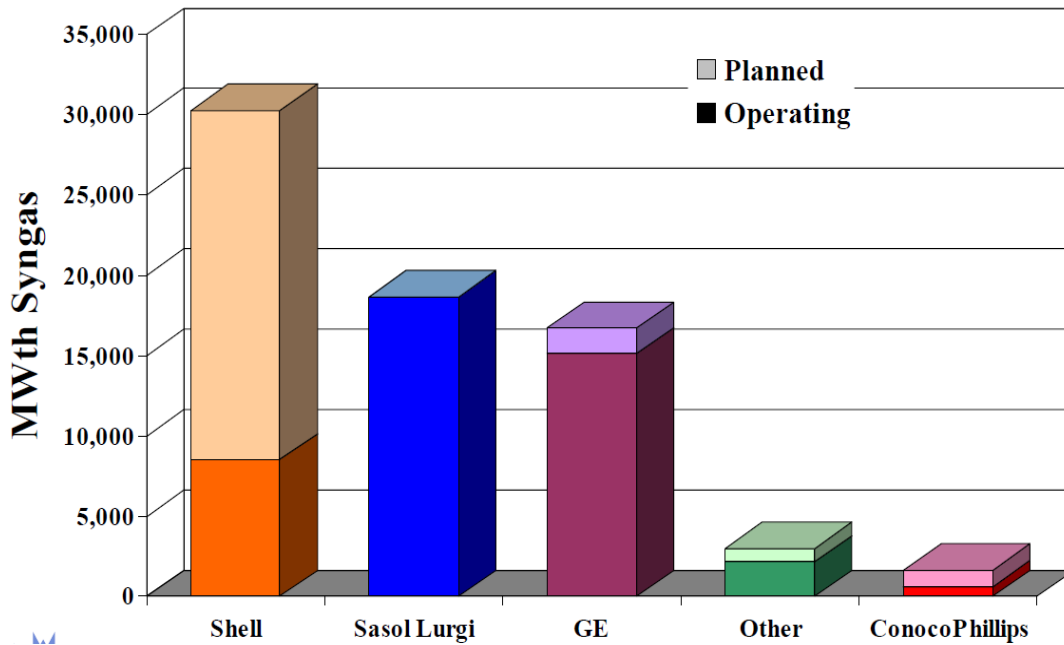


Figure 7: Gasification types by technology [15]

Table 5 illustrates the fate of GHG's produced from combustion versus gasification

Table 5: Environmental contrasts between a combustion and a gasification process [15]

	Combustion	Gasification
Sulphur converted to	SO ₂	H ₂ S
Sulphur capture	Flue gas scrubbers, boiler limestone injection	Absorbed in physical or chemical solvents
Sulphur disposal	Gypsum sold for wallboard	Sold as H ₂ SO ₄ or elemental S
Nitrogen converted to	NO _x	Traces of NH ₃ in syngas (syngas combustion produces low levels of NO _x)
NO_x control	Required (e.g., low-NO _x burners, staged combustion SCR/SNCR)	Currently not needed for IGCC (but tighter regs could require SCR)
C is converted to	CO ₂	Mostly CO in syngas
CO₂ control	Post-combustion removal from diluted stream	Pre-combustion removal from concentrated stream
Water requirements	Much more steam cycle cooling water needed	Some water needed for slurry, steam cycle and process needs

One of the major concerns however with the IGCC is that since it is a highly complex and integrated process, individual process/unit operations failures would result in major down times, thus reducing the overall plant availability and consequently its capacity factor. Major negative economic impacts are being observed due to the aforementioned problem. The following graph shows the availabilities reported from the IGCC units in operation. It can be seen from the graph that the assumed and previously proposed capacity factors for these power plants, in the range of 80% and higher, is not very realistic and several years has taken these technologies to even accomplish 80% availabilities. This underperformance significantly impacts the overall economics of the power plant [8].

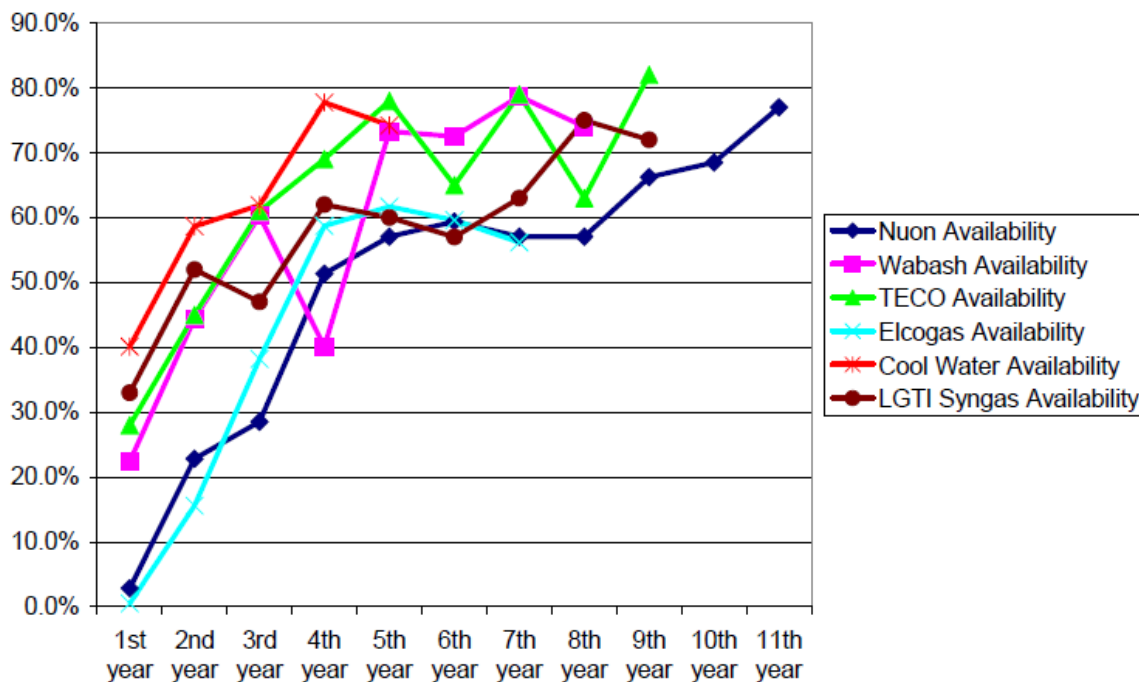


Figure 8: IGCC availability history (excluding operation on back up fuel) [18]

3.1.3 Oxy-fuel combustion with CO₂ capture

Oxy-fuel combustion employs the novel idea of the elimination of the non-reacting nitrogen component of air from the combustion gas thus reducing boiler and downstream cleanup processes. However, today's temperature limitations for boiler materials does not allow for

the combustion of coal with pure oxygen. In order to achieve reductions in the combustion temperature a portion of the flue gas, approximately $2/3$, is recycled and mixed with the pure oxygen stream entering the boiler. With the elimination of nitrogen from the combustion reaction the flue gas will contain mainly CO_2 and water vapour with minor amounts of impurities, thus minor additional purification is required to obtain a high purity CO_2 stream. The resulting high purity CO_2 stream is further purified in the cooling and compression stages of the CO_2 capture plant to arrive at the desired final CO_2 purities and pressures. Furthermore, NO_x removal processes are not required due to the absence of nitrogen. One of the major draw backs of this technology is however the expensive and power consuming air separation unit along with large quantities of power required for the CO_2 capture plant.

It should be noted that the only mature oxygen production technology with high enough oxygen production capabilities to date is the cryogenic air separation process. Large quantities of high quality power are required to drive the compressor shafts of the cryogenic air separation unit. However, there are various integration and heat recovery possibilities both from the air separation unit's compressors and from the CO_2 compression unit's operation. Currently both industry and academia are working towards improving boiler materials both in terms of temperature and acid gas resistances in order to enable elimination of flue gas desulfurization units and minimize the amount of recycle gas stream required. Also air separation technology providers are working on increasing their air separation units' (ASU) capacities so that a single train can provide the oxy-coal plant with its required amount of oxygen. Currently for a 500 MWe power plant, which is a common power plant size in Canada more than one ASU unit is required according to its oxygen consumption rates and ASU train capacities commercially available. The record holder for the largest ASU built is Air Liquide for a 2×4300 tonne/day unit built in South Africa [30]. A simplified overall process diagram of a typical oxy-combustion process with CO_2 capture is provided in Figure 9.

Pulverized Coal Oxycombustion

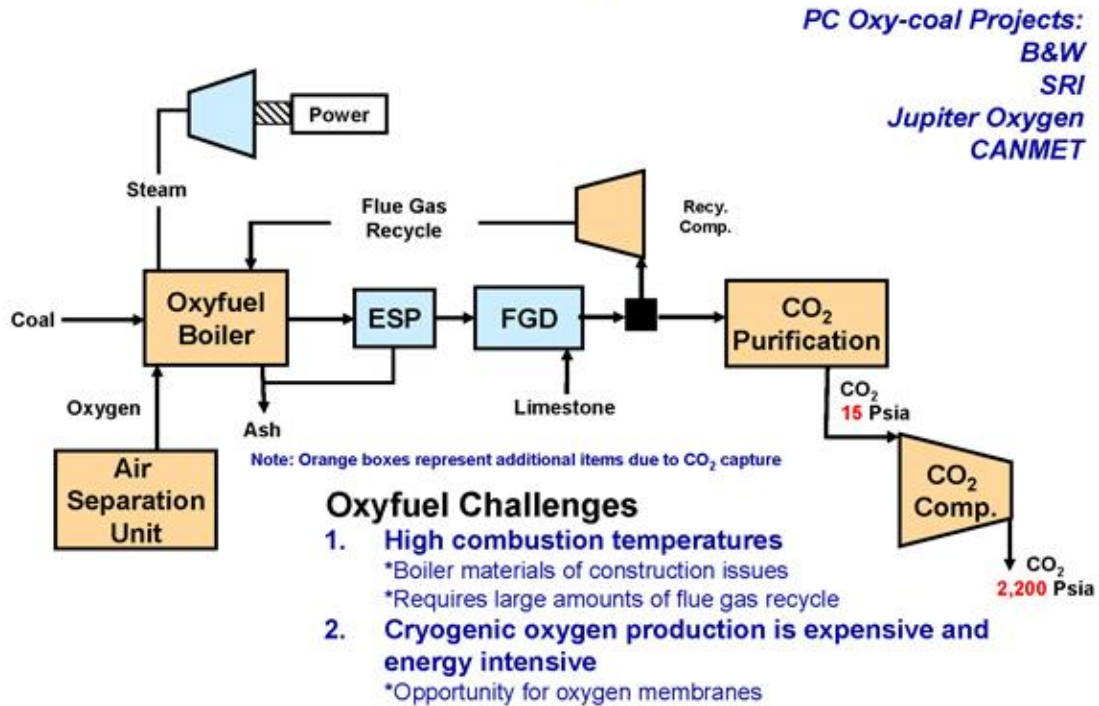


Figure 9: Oxy-fuel combustion [35]

3.2 Multi-technology assessment studies

Several major reports have assessed multiple technologies including post-combustion, pre-combustion and oxy-fuel processes in great detail. The results obtained from these studies are more reliable for comparison purposes due to the minimized variability introduced due to commonalities in the fundamental assumptions made. It should be noted that some of the case studies reported by the studies referenced below were eliminated due to their lower CO₂ purity requirements (<95%). This is due to the fact that lower purity assumptions eliminate possible markets, such as enhanced oil recovery (EOR) markets, in addition to reducing the number of options for storage sites. The results from these studies can be seen in Table 6. The cost of electricity (COE) for a certain technology might vary from study to study due to the differing fundamental assumptions employed; however the relative costs of the different technologies noted within a study is of great interest. It should be noted that all costs and

currencies were updated and adjusted to have a common basis of first quarter 2009 US\$ costs using chemical engineering plant cost indices (CEPCI) and exchange rates provided in the studies. Several issues need to be addressed before making any conclusions. In the first study performed by CCSD [11] it can be observed that the CO₂ capture rates are significantly different for the IGCC cases compared to the USC oxy-fuel case. It can be seen that oxy-fuel is the most economical technology choice for CO₂ capture when compared with the highest CO₂ capturing IGCC cases, even though they have lower than the recoveries reported for the oxy-fuel case. Looking at the second study performed by the CCPC [32] it can be seen that the lowest capture costs are reported for the IGCC Chevron Texaco case using Sub-bituminous coal, while the SC PC oxy-fuel is reported as the most expensive choice. However, there are some issues that need to be considered before drawing any conclusions, namely an assumption of full air-firing capacity for the oxy-fuel case which results in disregarding most of the equipment size reduction advantages normally attributed to the employment of oxy-coal combustion processes. Secondly, similar capacity factors are assumed and at high levels of 90% which is unrealistic for the IGCC as observed from Figure 8. It should be noted that it is generally believed that oxy-fuel combustion should have high capacity factors similar to or slightly less than their air-fired combustion counterparts, not their IGCC counterparts. Thus these inconsistencies might result in erroneous conclusions being drawn about the economics of the two technologies.

Table 6: Cost comparison of power generation technologies with CO₂ capture (costs in 2008 USD)

Study	Fuel type	Technology	Capacity Factor	COE €/kWh	MWe _{gross}	MWe _{net}	Capture (%)	Year
CCSD 2003[11]	Lignite	USC CCS Oxy-fuel	85.00%	4.46	598.60	450.00	95.00%	2002
	Lignite	IGCC Texaco MEA	80.00%	4.17	457.90	364.90	25.00%	
	Lignite	IGCC Texaco MEA	80.00%	5.87	405.40	271.90	75.00%	
	Lignite	IGCC Texaco Selexol	80.00%	5.42	-	-	75.00%	
CCPC 2004 [32]	Bituminous	IGCC Chevron Texaco	90.00%	10.08	594.00	444.50	87.00%	2003
	Sub-bitum.	IGCC Chevron Texaco	90.00%	9.15	629.00	436.80	92.00%	
	Lignite	IGCC Shell	90.00%	12.36	555.00	361.10	85.70%	
	Lignite	SC PC Amine absorption	90.00%	10.95	454.00	310.90	95.00%	
	Lignite	SC PC Oxy-fuel	90.00%	14.35	629.00	373.00	90.00%	
IEA 2006 [19] (Foster Wheeler)	German Brown	USC PC MEA	85.00%	6.82	932	761	85.00%	2005
	German Brown	USC CFB MEA	85.00%	6.75	1039.4	614.4	85.00%	
	German Brown	USC PCFB MEA	85.00%	7.02	763	688.4	85.00%	
	German Brown	USC PC Oxy-fuel	85.00%	6.91	816	741.3	93.00%	
	German Brown	IGCC Future Energy MDEA	85.00%	6.84	900.3	665.2	85.80%	
	German Brown	IGCC Shell MDEA	85.00%	7.51	868.7	628.8	85.20%	
	German Brown	IGCC FW MDEA	85.00%	7.13	900.5	686.6	82.90%	
DOE 2007 [17]	Bituminous	IGCC GEE Selexol	80.00%	11.60	744.96	555.68	90.00%	2007
	Bituminous	IGCC CoP E-GAS Selexol	80.00%	11.91	693.84	518.24	88.00%	
	Bituminous	IGCC Shell	80.00%	12.44	693.56	517.14	90.00%	
	Bituminous	SubC PC Amine absorption	85.00%	13.39	679.92	549.61	90.00%	
	Bituminous	SC PC Amine absorption	85.00%	12.94	663.45	546.00	90.00%	
DOE 2007 [16]	Bituminous	SC PC Amine absorption	85.00%	12.30	661.10	549.97	90.00%	2007 (Revised 2008)
	Bituminous	USC PC Amine absorption	85.00%	11.60	644.40	549.96	90.00%	
	Bituminous	SC PC Oxy-fuel	85.00%	11.56	785.90	550.01	92.90%	
	Bituminous	USC PC Oxy-fuel	85.00%	11.11	759.20	549.98	93.80%	
	Bituminous	SC PC Oxy-fuel w/ITM O2	85.00%	11.59	687.90	550.00	86.00%	
CCPC 2008 [9]	Sub-bitum.	ASC PC Oxy-fuel	85.00%	11.75	570.5	400.2	90.00%	2007
	Sub-bitum.	ASC PC Amine absorption	85.00%	11.87	480.5	391.3	87.00%	
	Bituminous	ASC PC Oxy-fuel	85.00%	12.74	568.1	413.2	89.00%	
	Bituminous	ASC PC Amine absorption	85.00%	11.79	490.7	409.9	88.00%	
	Lignite	ASC PC Oxy-fuel	85.00%	13.97	580	397.5	89.00%	
	Lignite	ASC PC Amine absorption	85.00%	12.95	479.2	382	87.00%	
	Lignite	IGCC Siemens Future Energy	85.00%	20.26	703	483	84.00%	
	Sub-bitum.	IGCC Slurry fed/Refractory	85.00%	15.20	620	481	74.00%	

Looking at the third study [19] it was reported that all electricity costs were roughly within 10% of each other. It can be seen that oxy-fuel combustion has the median cost of the reported electricity prices however with the consideration that it has significantly higher CO₂ recoveries. Assuming a CO₂ credit system being in place in the near future, this higher recovery would result in major attraction gains by this technology. In addition, sensitivity cases done by the authors indicated that for lower interest rate forecasts, oxy-fuel combustion becomes the most economical choice. The most economical technology choice was reported to be USC CFB (circulating fluidized bed) with MEA absorption in the study.

Looking at the fourth and the fifth studies conducted by DOE [16][17], which combined, produce the most comprehensive study for all currently available coal power production technologies and considering the studies had essentially the same financial basis and technology assessment methodology, it was seen that the three cheapest technology choices were all oxy-fuel based. The cheapest technologies being USC PC (pulverized coal) oxy-fuel combustion followed by SC PC oxy-fuel combustion followed by SC PC oxy-fuel combustion with ITM for oxygen production followed by USC PC amine absorption and the GEE IGCC with Selexol absorption processes.

The final study of consideration, the 6th study, was performed by CCPC [9]. This study is the second phase of investigation of the second study listed in Table 6. From the results reported by this study it was observed that oxy-fuel combustion and amine absorption had similar economics with one being superior to the other depending on the coal type considered. IGCC technologies had the highest costs of all technologies considered, however it should be noted that the IGCC technologies investigated were new generation ones which are not yet commercially available.

4 ECONOMICS

4.1 Financial modeling

In this study an in-house financial model has been developed both for the financing of an advanced super critical oxy-coal power plant with CO₂ capture and an ASC air fired coal plant without CO₂ capture capabilities. Being developed in Microsoft Excel, it is a robust and easy to use financing tool, which allows users to alter several parameters to obtain a more realistic assessment of their to-be-built power plant's finances.

The main outputs of the model are Levelized Costs of Electricity (LCOE) for both an air fired ASC without capture and an ASC oxy-coal power plant with capture capabilities along with a Levelized Avoidance Cost of CO₂ (LACOC). LACOC reflects the cost of reduction of CO₂ emissions by one unit, considering the amount of useful electrical output of the with capture plant remains the same as that of the power plant without capture [24]. There are several inputs into the model with reasonable default values in place. Some of the main components/parameters of the financial model are briefly discussed in sections 4.1.1 to 4.1.11.

4.1.1 Capacity factor

Capacity factor is defined as the ratio of the actual power output of the plant over a year and the output it would have had if it was operating at full capacity throughout the year. Capacity factors of most Canadian coal power plants are in the high 80's to low 90%'s range [7]. Oxy-fuel combustion is expected to have similar capacity factors to those of conventional coal power plants. However the employed capacity factor was 85% which is the suggested value by IEA GHG's economic assessment criteria [23]. It should also be noted that capacity factors higher than 85% are expected to be attainable in the long term for these plants. Sensitivity to capacity factor has been provided to assess the effect of uncertainties in this parameter which is presented in section 6.3.

4.1.2 Fuel consumption

The fuel consumption rate is obtained directly from a detailed boiler simulation provided by CANMET. The fuel chosen for this study is Saskatchewan lignite. As for the price of Saskatchewan lignite there are some uncertainties however an estimate for post 2013 reported by the CCPC (\$1.5/GJ) has been employed to obtain a \$/tonne price of coal, based on its heating value [9].

4.1.3 Escalation parameters

There are two escalation factors implemented in the model, one for fuel and the other for O&M costs, both of which are assumed to be 2% per year. This number was used to escalate all costs from the Commercial Operation start Date (COD) of the power plant. No escalation in prices is included for the construction period.

4.1.4 Financial profile

During the construction period a short term financing profile of 100% debt was assumed, along with an interest rate of 1% higher than the long-term debt financing interest rates employed. The investment profile was then changed to a long term financing comprised of both debt and equity components from the power plant's COD. The long term debt financing interest rate was set at 7% as the default value. Sensitivity of the economics to long term debt interest rate is provided in section 6.4.

4.1.5 Construction duration

The duration of construction commonly varies from 3 to 6 years, however, since the oxy-fuel combustion process is more complex than a conventional air fired coal power plant, duration of 4 years was employed as the default value for the ASC oxy-coal plant and a duration of 3 years was assumed for the ASC air fired case. The tool however allows for construction durations of up to 6 years always maintaining the construction duration of the air case one year less than that of the oxy-fuel process.

4.1.6 Taxation

The taxation rate was obtained from KPMG's forecast for post 2012 taxation rates. It is a combination of the federal and provincial taxation rates. The default rates are those corresponding to Ontario's but the user can alter both taxation rates to obtain case specific results based on the power plant's location [28].

4.1.7 Depreciation and capital cost allowance

Using appropriate depreciation models is very important in evaluating worth of assets of a company at any point in time. However, for evaluation of to-be-built power plants where the salvage value of the power plant does not have a significant impact on the economics, the importance of depreciation calculations arise in tax calculations. For taxation purposes the assets depreciate according to a Capital Cost Allowance rate (CCA) prescribed by the government [20]. Two CCA categories were employed in this study. A declining-balance depreciation model as typically employed in Canada along with the implementation of the half year rule was built into the financial model.

It should be noted that there are usually incentives provided by the government in the form of higher capital cost allowances to make clean/green technologies more affordable. Currently for CCS projects there are not any significant incentives in place however if they are put in place in the form of higher CCA rates CCS projects will become a more economical solution to the high GHG emission problem.

4.1.8 Distribution of investment requirements during the construction period

The distribution of capital requirements during the course of the construction is developed based on reported distributions developed by major vendors in the literature [14][19]. Investment schedules commonly follow a normal bell shaped curve that is skewed towards the commercial operation start date (COD) in terms of capital requirements. A sensitivity analysis to the duration of construction is provided in section 6.6.

4.1.9 Levelized cost of electricity

The levelized cost of electricity is defined as the constant cost of electricity required to balance all the costs and revenues over the life of the power plant. It is a constant cost which is in essence a slight over estimation of the cost of electricity at the beginning years of the plant's operation and an underestimation of the electricity costs towards the end of the economic life of the plant. The mathematical equation involved in LCOE calculation can be presented as follows:

$$LCOE = \frac{NPV(Operating+Debt+Taxes+Equity)}{(CF)(8760)(MW)} \quad \text{Equation 1}$$

Where:

- LCOE = iteratively determined constant cost of electricity in \$/MWh_{net}
- NPV = net present value of the sum of all costs' yearly sums over the economic life of the plant discounted at the rate of the cost of equity; Operating = fuel + O&M + transmission costs (5% of yearly energy revenues for transmitting the electricity) including cost escalation (\$/year, not constant)
- Debt = Annual debt principal and interest repayment (\$/year, not constant)
- Taxes = Annual taxes repayments on taxable income (\$/year, not constant)
- Equity = Equity principal repayment (\$/year, not constant)
- CF = plant capacity factor
- 8760 = Total number hours in a year
- MW = net MW electric output of the plant

4.1.10 Cost of CO₂ avoided

The cost of CO₂ avoided is calculated using the levelized cost of electricity and the emissions of a power plant with CO₂ capture and one without CO₂ capture. The choice of power plant used for the without capture case is an ASC air fired coal power plant with similar power outputs to the ASC oxy-coal power plant with capture capabilities.

$$LACOC = \frac{(LCOE)_{w/ \text{ capture}} - (LCOE)_{w/o \text{ capture}}}{(Emissions)_{w/o \text{ capture}} - (Emissions)_{\text{capture}}} \quad \text{Equation 2}$$

Where:

- LACOC=levelized avoidance cost of CO₂ (\$/tonne)
- LCOE_{w/ capture} = levelized cost of electricity (\$/MWh_{net}) with capture
- LCOE_{w/o capture}= levelized cost of electricity (\$/MWh_{net}) without capture
- Emissions_{w/o capture}=Emission intensity of a power plant without capture (tonnes/MWh_{net})
- Emissions_{w/ capture}=Emission intensity of a power plant with capture(tonnes/MWh_{net})

The values for the two emission intensities both for without and with capture cases were adopted from the CCPC phase II summary report, since a similar power plant process scheme and fuel is used for the CCPC study to the process simulated in-house.

4.1.11 Summary of assumptions

A summary of the important financial assumptions can be observed in Table 7.

Table 7. Economic assumptions

Parameter	Default values employed
Capacity factor (%)	85
Economic life of plant (years)	30
Construction start date (year)	2010
Construction duration (years)	4 (3 years for the air fired case)
Short-term interest on debt during construction (100% debt)	8
Debt ratio (fraction)	0.5
Interest on debt (%)	7
Debt repayment period (years)	30
Return on equity (%)	15
Lignite price (\$/tonne)*	23.41
Fuel cost escalation (%)	2
Other cost escalation (%)	2
Combined capital cost allowance rate employed (%) **	5
Taxation (Combined federal & provincial) (%)***	29

*Based on 1.5 \$/GJ lignite price reported for post 2013 in the CCPC phase II summary report [9]

**Based on two different CCA rates of 8% and 4% pertaining to different power plant areas

***KPMG's forecast for post 2012 taxation rates for Ontario [28]

4.2 Cost estimation methodology

Different cost estimation methodologies were employed for the evaluation of the different process areas. A list of methodologies employed in performing the capital cost estimations are:

- Vendor quotations/confirmations (when available);
- Costing software/tools (ex. Aspen IPE);
- Scaled estimates from previous design/build projects and references; and,
- Costing correlations from plant design/costing references.

As for the scaled estimates from previous design/build projects and references there were two constraints imposed on the selection of sources to be employed:

1. Reference studies would have to be published post 2005, the reason being that, it is believed that costs reported prior to the year 2005 are not considered as valid and reliable for an emerging technology and with the dramatic recent changes in prices [7].
2. The numbers provided by the source study must be based on detailed process simulations for which cost estimations were provided using either vendor quotations or developed by globally accredited contracting companies.

After considering the above constraints three studies were found suitable for implementation in the capital cost estimations [14][16][19], and the costs reported by these studies were further updated, scaled and adjusted to develop a common basis for this study's cost estimates.

4.2.1 Capital cost estimation

The power plant was segregated into the following main process areas, for each of which different cost estimation technique/s was/were employed as listed in Table 8.

Table 8: Cost estimation methodologies employed for the ASC oxy-coal with CO₂ capture

Process Area	Scaled estimates from previous design/build project & references	Costing software	Vendor Quotations/ Confirmation
Coal/sorbent preparation, ash & spent sorbent handling	√		
Air separation unit	√		√
Boiler and accessories	√		
Balance of plant	√		
Flue gas cleanup		√	
CO ₂ capture & compression unit		√	√
Utilities & off sites	√		

It should be noted that for the air-fired ASC coal power plant the only costing tool employed for evaluating all process areas was scaled estimates from references. Furthermore when using scaled estimates for process areas' cost evaluations, since each study has employed slightly different codes of account or area/facility account structure, different process components' costs were included in different process areas; thus, when possible, adjustments to the allocation of these process components were made.

The costs included by these studies employed different breakdowns for coming up with installed costs, as well; however the cost components included were very similar, thus acceptable. In addition to installed costs three extra component costs were included in the studies presented in the IEA GHG, reports 2005/09 and 2006/01 which were contingencies, fees and owner's costs [14][19].

Contingencies are determined based on the level of detail of the study performed thus should not be meddled with, and the estimates provided for this parameter were deemed appropriate. The only exception was the extra contingency component (process contingency)

implemented by the U.S. department of energy (DOE) for the ASC oxy-boiler costing strictly for the relative immaturity of ASC oxy-fuel combustion technology and for the ASC air-fired technology[16]. These contingencies were added to the other two studies since they had been carried out even earlier.

Furthermore it should be noted that for the flue gas cleanup process and CO₂ capture and compression unit contingencies for the ASC oxy-coal case were recommended by ASPEN IPE based on the nature of the process, and the degree of complexity.

As for the owner's costs, DOE had excluded all owner's costs which are to include:

- permits and licensing (other than construction permits);
- land acquisition / rights of way costs;
- economic development;
- project development costs;
- legal fees; and
- owner's Engineering / Project and Construction Management Staff.

A cost and performance baseline study by the DOE estimates these costs to fall in the range of 15-25% of the total plant costs [17].

The IEA GHG report 2006/01 [19] had implemented a 5% owner's costs devoted to land purchases and surveys; however this number was considered to be an underestimation as the other two studies, which had higher accuracies, both suggest numbers in the 15-25% range of capital costs [17][14].

In order to overcome this issue 20% of installed costs was implemented as suggested by the more detailed IEA GHG report 2005/09 [14] for all areas except for Air separation unit and the CO₂ capture and compression unit for which a 5% value was employed by both IEA studies [14][19]. As for the DOE's ASC oxy-fuel and ASC air fired studies [16] since the

installed cost's break down was slightly different, the owner's costs calculated for the other studies were expressed on a \$/MW_{net} basis and this was implemented as an addition to the capital cost estimates of the DOE's reported costs.

4.2.1.1 Correcting for the size of each process area's capacities

Typically for correcting for the size of process equipment, the “0.6 rule” is employed which can be expressed in a mathematical form as follows [43][37]:

$$C_{P,v,r} = C_{P,u,r} \times \left(\frac{v}{u}\right)^a \quad \text{Equation 3}$$

Where:

- $C_{P,v,r}$ is the purchase price of the equipment in question which has a size or capacity of v in the year r
- $C_{P,u,r}$ is the purchase price of the same type of equipment in the same year but of capacity or size u.

The generic exponent which applies to most unit operations is $a=0.6$ in the above equation hence commonly referred to as the “0.6 rule” [42]. However, the accuracy of updated costs using this method could decrease for certain processes due to the exponent of 0.6 being inappropriate; thus, when available, more process specific coefficients were applied.

Integrated Environmental Control Model (IECM) developed for the DOE provides more process area specific coefficients to be employed in place of the value of 0.6 [5]. However it should be noted that the application of these coefficients requires two assumptions to be made. IECM had calculated these coefficients based on cost of each process area for a number of subcritical power plants burning sub-bituminous coal. Thus it was assumed that firstly the sub-bituminous cases' cost numbers could be applied to power plants burning lignite coal and secondly these coefficients which were developed for subcritical power plants were applicable for power plants with more advanced steam cycles.

For process areas such as the air separation unit and flue gas cleanup since area specific coefficients were unavailable the coefficient of 0.6 was employed in the calculations.

4.2.1.2 Correcting for the studies' estimate dates

For updating the costs reported by each study, Chemical Engineering Plant Cost Index (CEPCI) was employed. The mathematical equation employed for updating the costs was as follows:

$$C_{P,v,s} = C_{P,v,r} \left(\frac{I_s}{I_r} \right) \quad \text{Equation 4}$$

Where:

- $C_{P,v,s}$ is the updated cost in time s
- $C_{P,v,r}$ is the cost estimated in time r
- I_s is the current cost index
- I_r is the cost index at time r

It should be noted that large variations in the CEPCI was observed in the past year. The volatile global economic situation is partially responsible for these fluctuations. It is believed that the cost indices being reported are temporary low values due to the current economic downturn. Thus it was considered appropriate to average the CEPCI's for the past twelve most current months. Fluctuations were observed from 511.8 to 619.3 which result in large variations in the capital cost estimations when employing scaled reference estimates. Thus the average of the reported highest and lowest indices from May 2008 to April 2009 was employed which was equal to 565.55. A sensitivity analysis to the CEPCI is provided in section 6.1 indicating its effects on the project's economics.

4.3 Capital cost estimation results

A description of the process evaluated is provided in section 4.3.1. Further notes on capital cost estimations and the results obtained for each of the process areas are summarized in this section. Each sub-section has a graph of the area's capital costs after being scaled and adjusted from the references along with the final costs calculated for this study and its respective accuracy range, expressed in up to date US dollars.

4.3.1 Plant description

Cost estimations were developed for an advanced supercritical pulverized coal oxy-fuel power plant with CO₂ capture capabilities, using the cost estimation methodologies summarized in Table 8. The power plant evaluated is assumed to have a 500 MW gross capacity with a net electrical output of 380.5. Also cost estimations were developed for a 380.5 MW_{net} air fired advanced supercritical coal fired power plant without CO₂ capture capabilities strictly from scaled references conforming to the prescribed requirements detailed in section 4.2. A summary diagram of the different process areas of the oxy-fuel power plant can be observed in Figure 10.

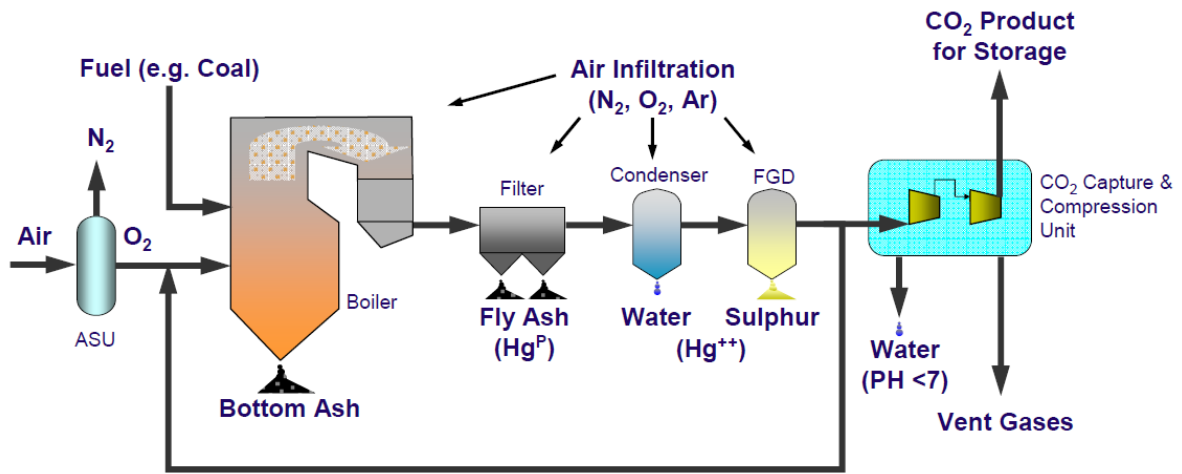


Figure 10: ASC oxyfuel combustion with CO₂ capture [48]

For the case of ASC oxy-fuel, an oxygen purity of 95% (volume basis) was deemed necessary being produced by 2*4550 tonne/day cryogenic air separation units having a three column design as shown in Figure 11.

A dry and desulphurized recycle stream configuration was deemed appropriate. Flue gas clean up included a cold ESP and a wet FGD. Two configurations were evaluated for the recycle stream with one being placed before the FGD and the other downstream of the FGD.

However in order to avoid corrosion issues the latter was selected as the configuration of choice.

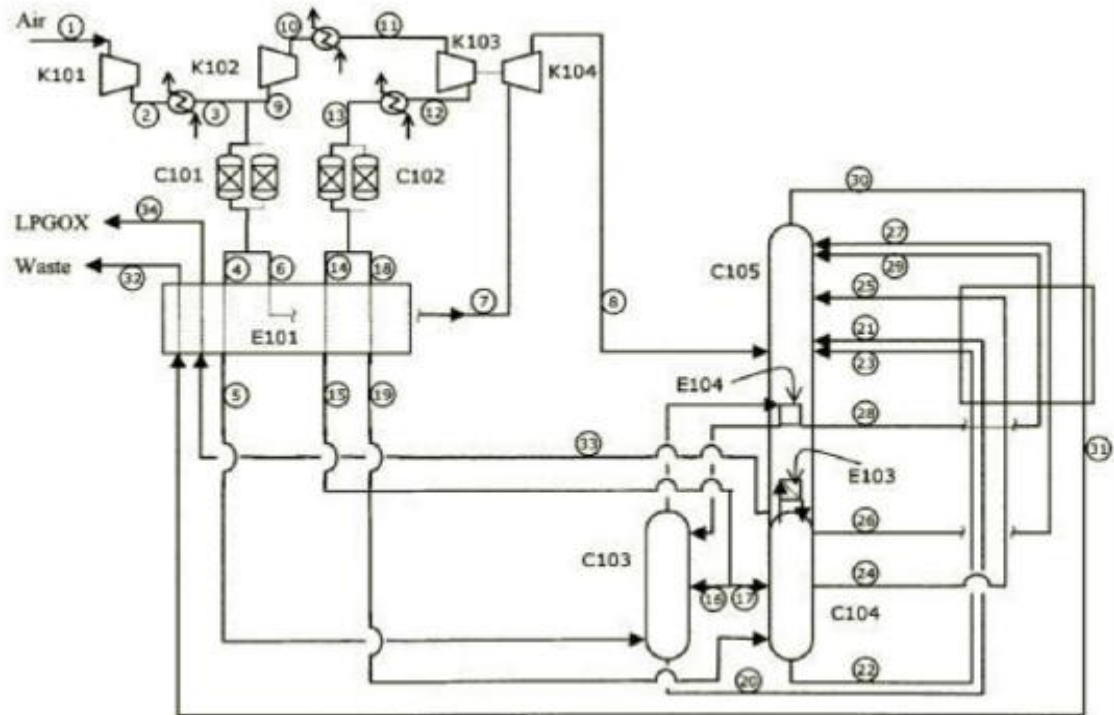


Figure 11: Three column design cryogenic air separation unit.[2]

A CO₂ recovery of 90%+ was achievable with this CO₂ capture unit simulated at 95%+ purity. A diagram of the CO₂ capture plant is provided in Figure 12.

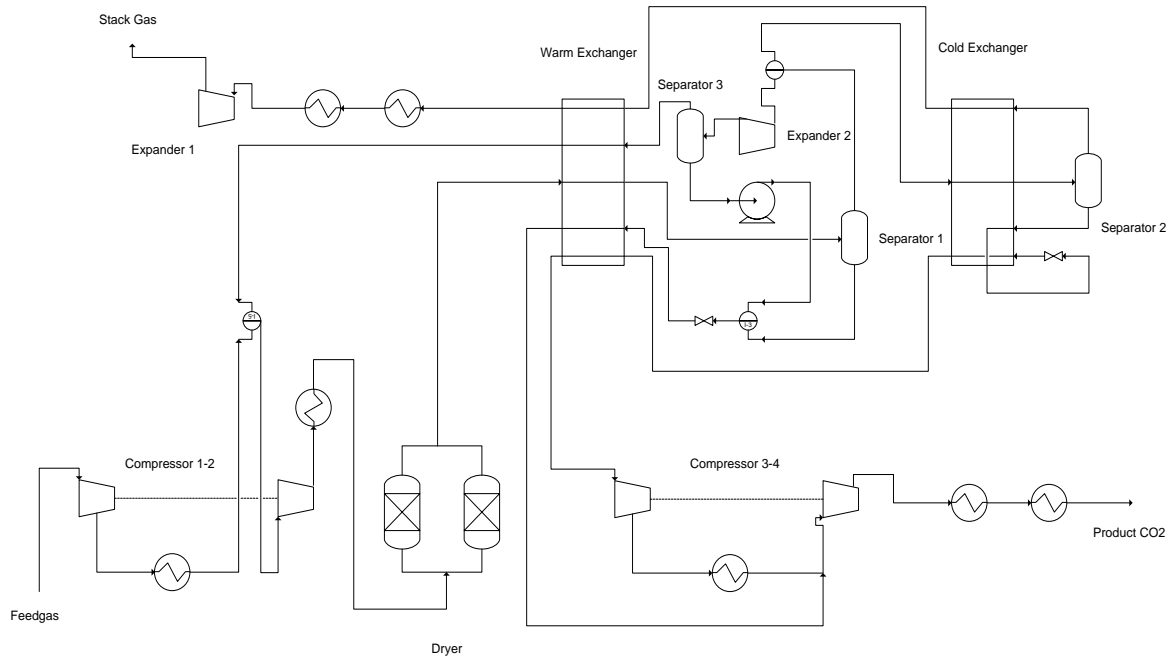


Figure 12: CO₂ capture and compression unit [47]

4.3.2 Coal/sorbent preparation, ash & spent sorbent handling

Three section costs, provided by the recent study conducted by the DOE [16], were combined to provide an estimate for the assumed scope of coal/sorbent, ash & spent sorbent handling process area, namely:

- Coal handling system;
- Coal prep & feed systems; and
- Ash/spent sorbent handling system.

Similarly, the IEA GHG report 2006/01 had provided two separate costs for coal preparation and coal drying processes which were combined in this assessment as well [19].

Furthermore it was noted that the two estimates provided by the two IEA GHG studies [14][19] did not include any sorbent prep, feed and spent sorbent handling section since their

designs did not employ an FGD unit. Thus, an estimate of these units on a $\$/\text{MWhr}_{\text{net}}$ basis was obtained from the detailed cost estimates provided by the DOE's study [16] and was added to the two studies to create a common process basis for the case of ASC oxy-fuel power plant. A summary of adjusted and corrected costs with the final estimate calculated is provided in Figure 13 for the ASC oxy-fuel case. It can be observed that there is a variation present in the reported costs. This variation is partially due to the difference in process schemes assumed, for instance how the drying of coal is carried out. However, it should be noted that since there was no detailed simulation of this process area provided in this study an average of the cost estimates was deemed appropriate.

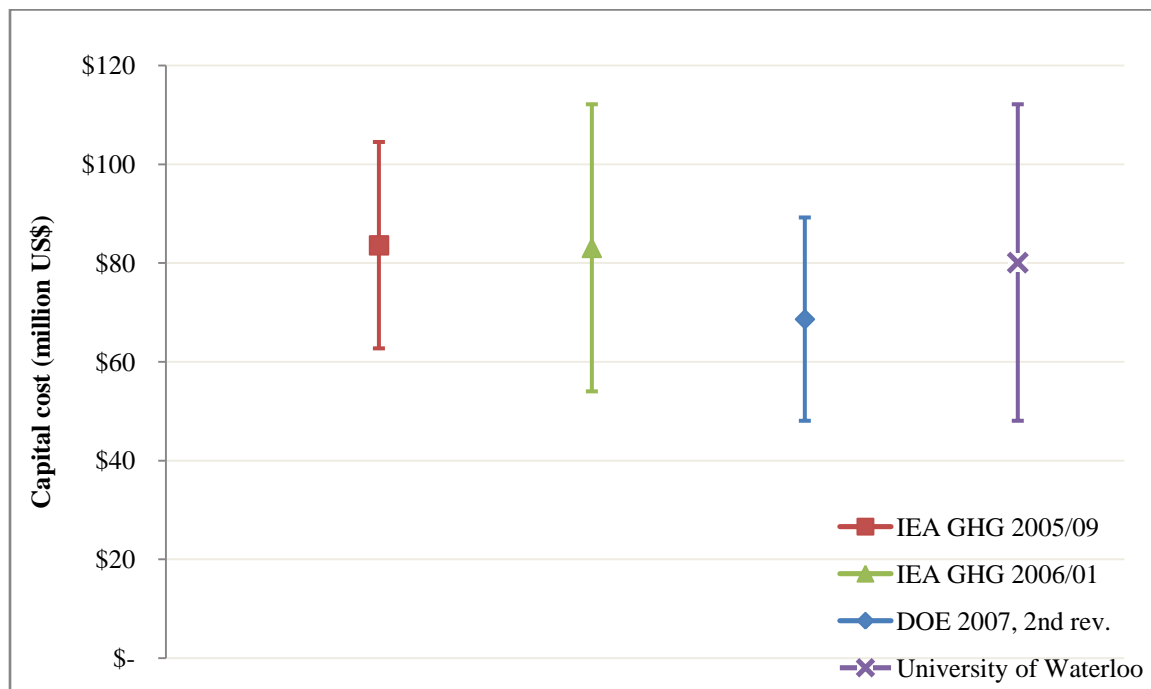


Figure 13: Coal/sorbent preparation, ash & spent sorbent handling costs for a 380.5 MW_{net} ASC oxy-coal plant with CO₂ capture

4.3.3 Air separation unit

For the air separation unit the three references were scaled and adjusted to obtain an estimate of the capital cost. It should be noted that a large variation in costs was observed. Air separation technologies have gained popularity in the recent years thus great attention has

been dedicated to optimization and integration of these processes, which naturally also result in cost reductions. Older studies are expected to have higher reported costs, as can be seen in Figure 14, where the more outdated the cost estimate the higher its reported value. In addition different technology providers have slightly different process schemes with varying capital and O&M costs. A recent paper presented at the GHGT9 conference briefly discusses some of these differences [2]. A vendor confirmation value was also obtained, which was not included in calculating the costs for this study however it is included as an extra reference number presented in Figure 14.

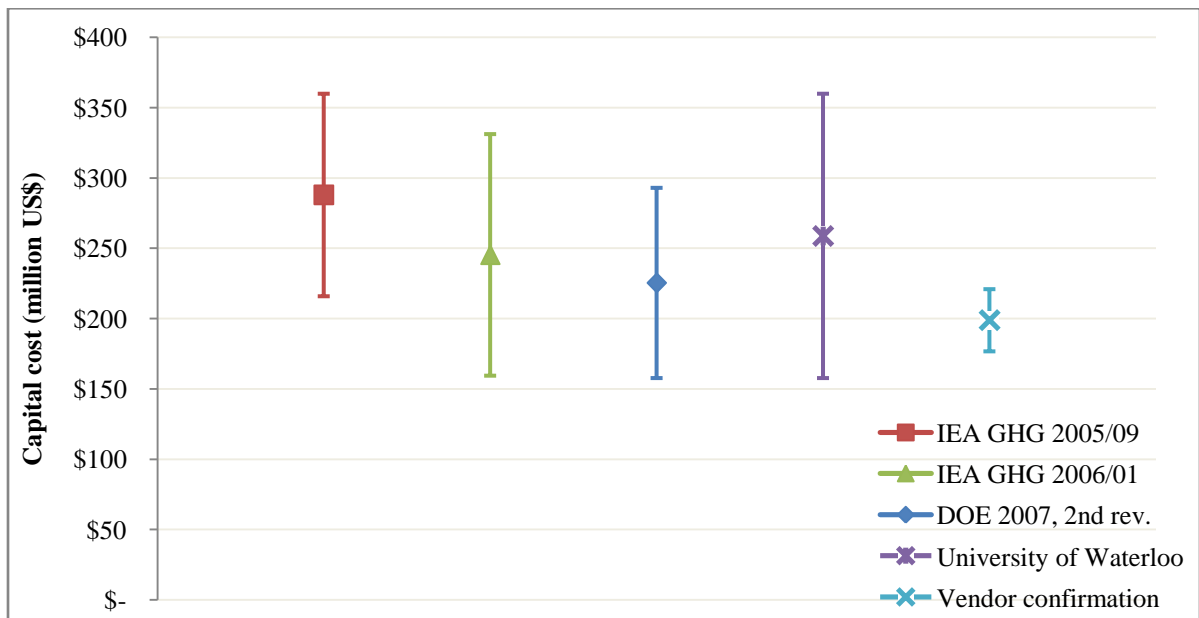


Figure 14: Air separation unit's capital costs for a 380.5 MW_{net} ASC oxy-coal plant with CO₂ capture

4.3.4 Boiler and accessories

The most recent study by the DOE [16] had included an extra 15% process contingency for the boiler strictly to take into account the maturity of the ASC oxy-fuel technology and an extra 10% for the ASC air-fired case to take into account the maturity of the ASC air-fired technology. Since there was no extra contingencies included dealing with the maturity of the technology by the other two source studies [14][19] an extra 15% and 10% of installed costs, of contingency was added, since they were prepared about two to three years prior to the

study conducted by the DOE. As far as the different process components included in the estimate are concerned the cost of air separation unit was included in the boiler costing reported by the DOE study which was separated. Also the boiler costs reported by the IEA studies include the cost of electrostatic precipitator thus they should slightly overestimate the cost of their boiler [14][19]. However it is observed that these two studies in fact report lower boiler costs. This discrepancy is attributed to the state of knowledge since boiler costing and operation in the oxy-combustion mode is unknown at a commercial stage. In these cases typically more confidence is given to the more recent designs presented. The boiler and accessories capital costs are shown in Figure 15.

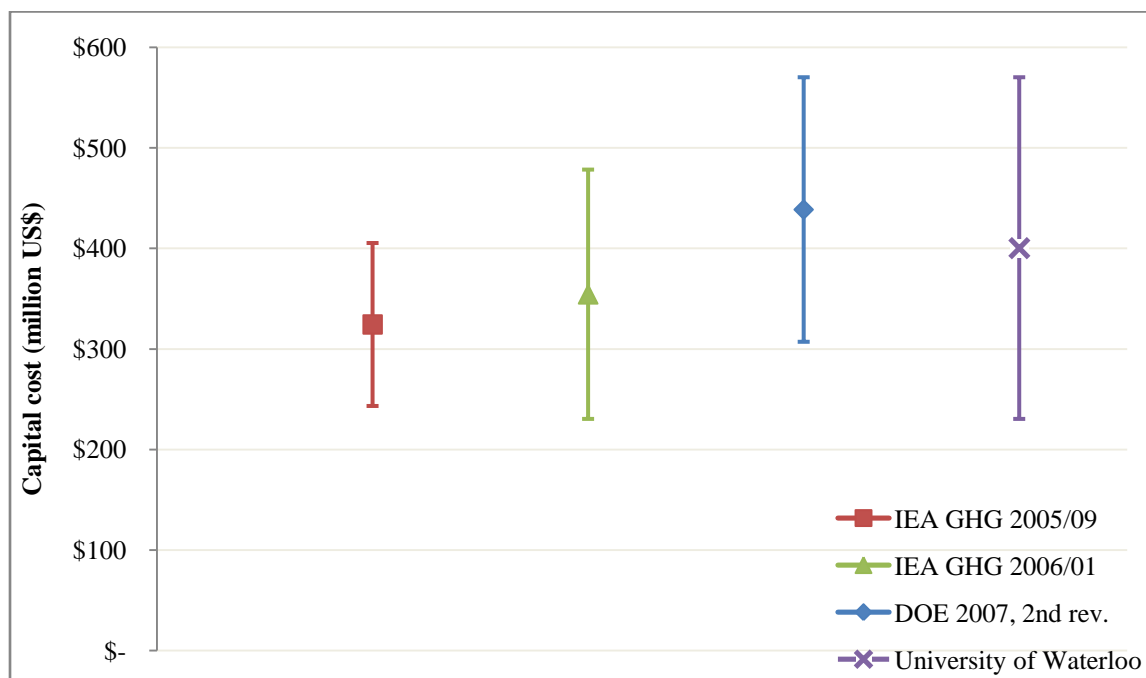


Figure 15: Boiler & accessories' capital costs for a 380.5 MW_{net} ASC oxy-coal plant with CO₂ capture

4.3.5 Balance of plant (BOP)

For the balance of plant costing the source studies were scaled and updated and the results can be seen in the figure below for the ASC oxy-fuel with CO₂ capture case. It should be noted that for the study conducted by the DOE the two process areas', feed water & misc. BOP systems and steam turbine generator costs were combined into one cost in order to

create a common basis for all three studies [14][16][19]. The BOP costs are presented in Figure 16.

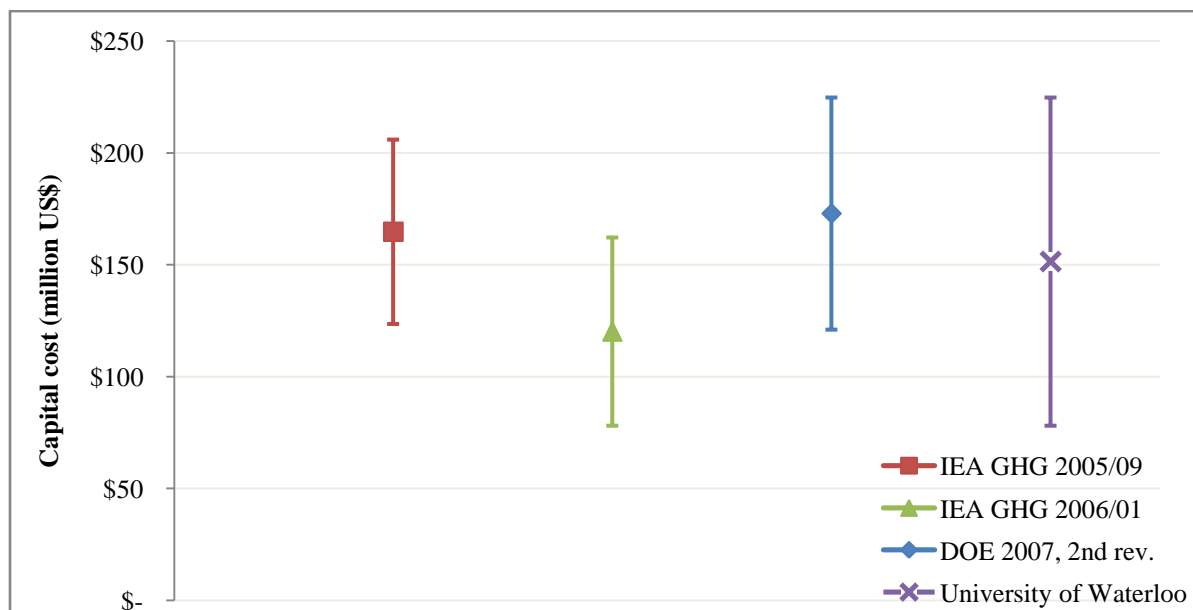


Figure 16: Balance of plant's capital costs for a 380.5 MW_{net} ASC oxy-coal plant with CO₂ capture

4.3.6 Flue gas clean up

Two configurations were considered in the estimation of costs for the flue gas cleanup process of the ASC oxy-fuel with CO₂ capture case. Aspen IPE was employed to provide a cost estimate of these units excluding the FGD unit for which the 0.6 rule was applied to the costs provided in the DOE's study [16]. From the two configurations the more expensive and conservative design was selected. The differences between the two configurations were the following:

1. FGD past recycle: after the ESP a gas/gas heat exchanger (HR1) followed by flue gas blowers was simulated. The flue gas exiting the blowers then passed through a gas/liquid heat exchanger (HR2) heating a condensed steam stream before being recycled back to the HR1 as the cooling medium. Roughly one third of the flue gases were at this point directed to the FGD with the remaining being recycled. In this

design the recycled flue gases are not desulphurized thus corrosion issues would need to be addressed.

2. FGD prior to recycle: the HR1 heat exchanger in the previous case is eliminated in this design and the FGD is assumed able to cool down the gases and carry HR1's cooling load. In this design all flue gas is desulphurized thus less corrosion concerns are present, however it is more expensive. The costs associated with this design were included in the flue gas cleanup costing.

A summary of the flue gas clean up costs calculated and adjusted can be seen in Figure 17.

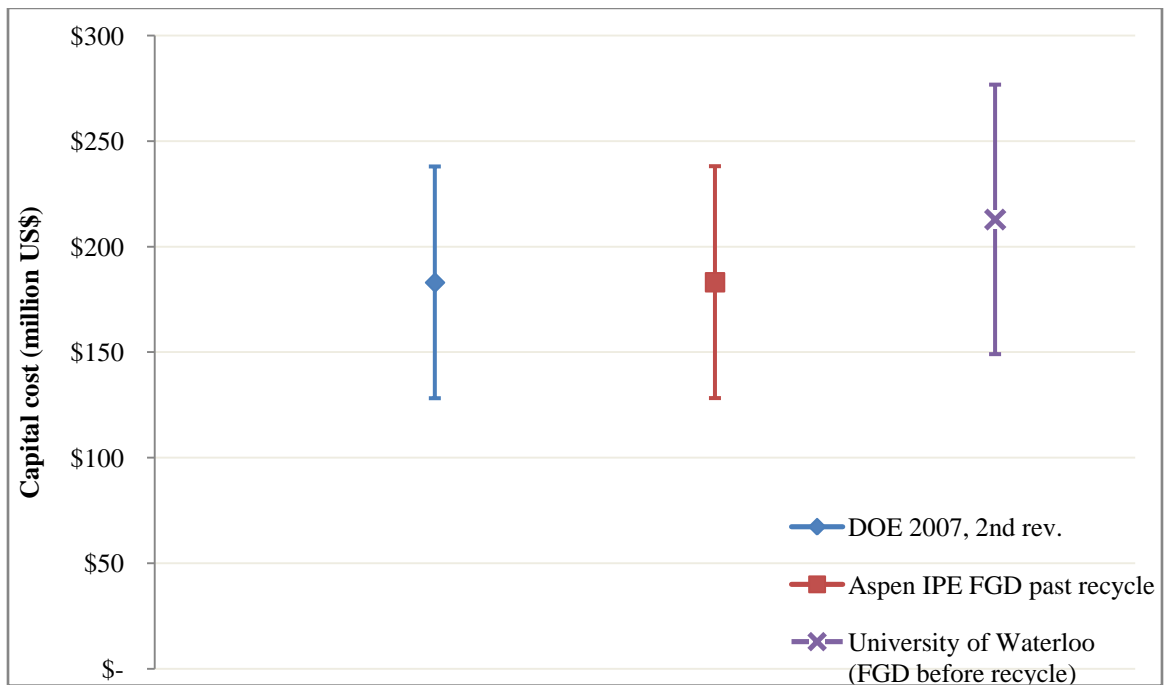


Figure 17: Flue gas cleanup unit's capital costs for a 380.5 MW_{net} ASC oxy-coal plant with CO₂ capture

4.3.7 CO₂ capture and compression unit

The cost of the CO₂ capture and compression unit was estimated using detailed designs and simulations using ASPEN plus and ASPEN IPE for providing accurate cost estimates. For the costing of Aluminum brazed heat exchangers and the molecular sieves used for exchanging heat and drying the flue gas respectively vendor quotations were obtained to

provide accurate estimations of the costs [29][38]. A summary of the costs obtained/calculated can be observed in Figure 18.

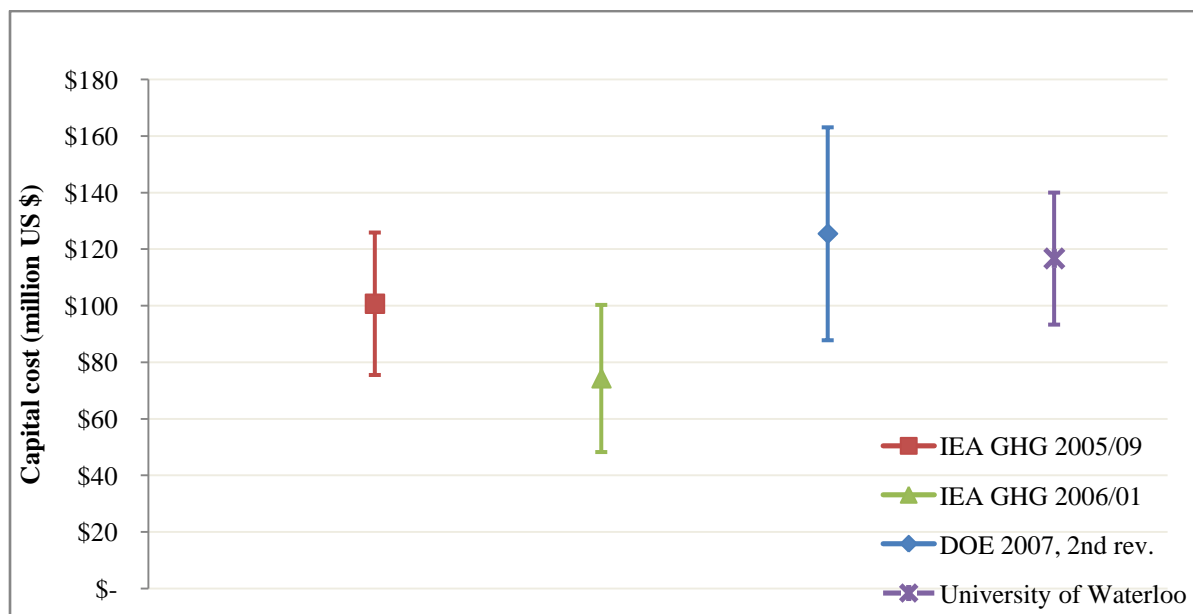


Figure 18: Capture and compression unit's capital costs for a 380.5 MW_{net} ASC oxy-coal plant with CO₂ capture

4.3.8 Utilities and off sites

Utilities and off sites were estimated from the reference studies. As for the DOE's study the costs for the cooling water system, accessory electric plant, instrumentation & control, improvements to site and building and structures' costs were combined to create a common basis of cost estimation for the studies used [14][16][19]. The costs are provided in Figure 19.

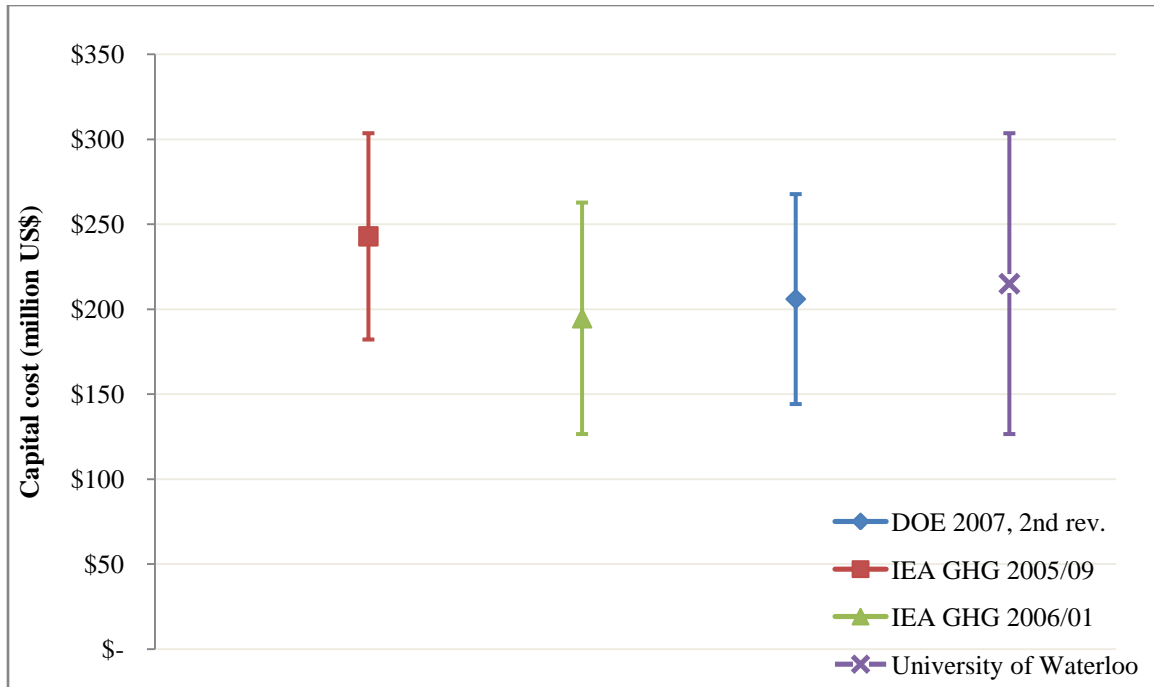


Figure 19: Utilities and off sites' capital costs for a 380.5 MW_{net} ASC oxy-coal plant with CO₂ capture

4.3.9 Summary of capital costs and cost estimation methodology

A summary of the capital costs is provided for the overall plant costs as well as the process areas along with overall plant cost estimates reported by the three reference studies in Figure 20 for the ASC oxy-coal with CO₂ capture case. As an extra source of comparison a cost estimate was developed using the integrated environmental control model (IECM). IECM is commonly employed as a rough cost estimation tool for which accuracies in the range of 30 to 50% is expected. The error bars indicated in Figure 20 for the IECM cost estimate correspond to +/- 50% error. It should be noted that the price of coal and its composition along with any other adjustable financial and process parameter were inputted into IECM to resemble those employed for the cost estimations carried out in this study. For estimating the different process areas accuracy bounds reported in reference studies' the claimed accuracies were used developing minimum and maximum costs for each of the process areas for each of the studies summarized in sections 4.3.2 to 4.3.8. The accuracy range of the process areas' cost estimates for this study was developed using the minimum of these minimums

from the three source studies along with the maximum of their maximums respectively as the minimum and the maximum of the cost estimates.

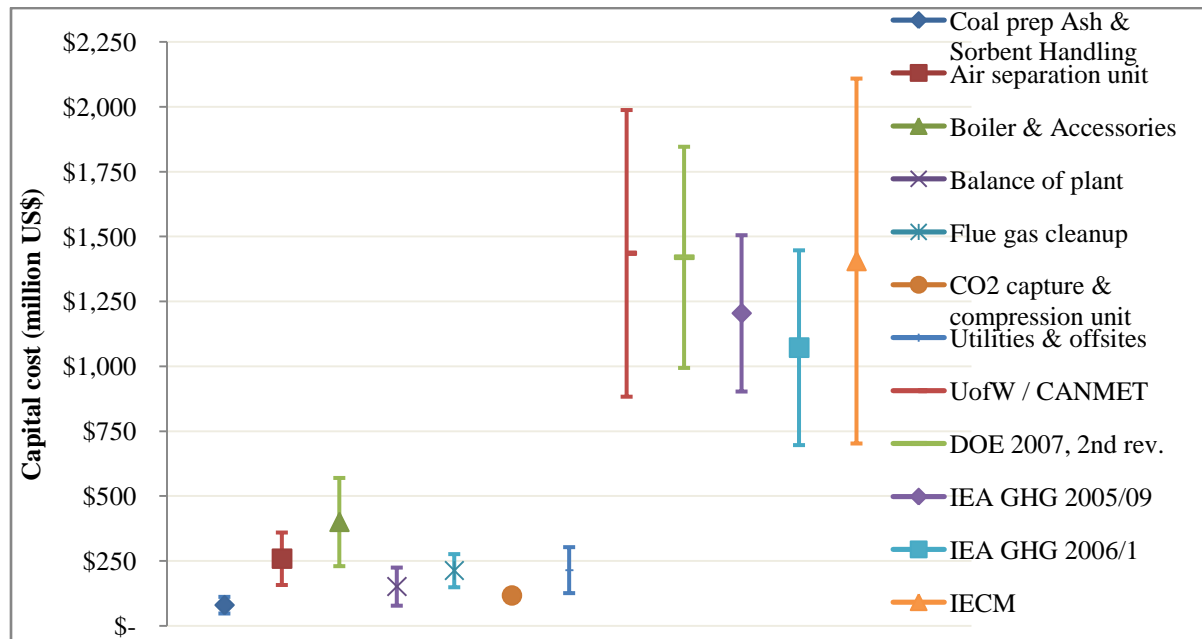


Figure 20: Summary of capital costs for a 380.5 MW_{net} ASC oxy-coal plant with CO₂ capture

Furthermore the average of these two values was employed as the estimate of the cost of the process area. This approach is a very conservative yet effective method. With the lack of case specific details such as location of the plant and the absence of detailed vendor quotations guaranteeing smaller ranges for cost estimates is infeasible. Similarly for the CO₂ CCU and the flue gas cleanup the level of accuracy reported although using costing software (which potentially can provide estimates with very high accuracies) was not very high, since case specific information is required in order to be able to claim higher accuracies. Capital cost estimates were also developed for the air fired case without capture, from scaled references as well for which the results are summarized along with those pertaining to the ASC oxy-coal with capture case in Table 9.

Table 9: Summary of capital costs

	ASC oxy-coal with CO ₂ capture		Air fired ASC without CO ₂ capture	
Process area	Cost estimated (million US \$)	Accuracy (+/- %)	Cost estimated (million US \$)	Accuracy (+/- %)
Coal prep ash & sorbent	\$ 80	40%	\$ 57	35%
ASU	\$ 259	39%	\$ -	0%
Boiler	\$ 400	42%	\$ 313	39%
BOP	\$ 151	48%	\$ 152	31%
Flue gas cleanup	\$ 213	30%	\$ 115	35%
CCU	\$ 117	20%	\$ -	0%
Utilities & off sites	\$ 215	41%	\$ 213	47%
Overall capital costs	\$ 1,435	38%	\$ 850	39%
Overall capital costs DOE 2007, 2 nd rev.	\$ 1,420	30%	\$ 822	30%
Overall capital costs IEA GHG report 2005/09	\$ 1,204	25%	\$ 755	25%
Overall capital costs IEA GHG report 2006/01	\$ 1,072	35%	N/A	N/A
IECM	\$ 1,406	50%	\$ 822	50%

It was also observed that the largest accuracy boundaries are those pertaining to the boiler & accessories and the balance of plant cost estimates for the case of ASC oxy-coal with CO₂ capture, whereas for the case of air fired coal without capture the largest error in estimates were those of the boiler and the utilities and off sites process areas.

4.4 Operations & maintenance costs (O&M)

There are two portions to the Operation and Maintenance (O&M) costs namely fixed and variable O&M costs. Fixed costs entail those that are not a function of the power output and merely they will occur no matter what the output of the power plant is. Variable costs on the other hand are a direct function of the power plant's output and hours of operation. Some costs can be included in either; however, depending on the level of detail of the study a decision needs to be made as to which category each cost should be allocated in. For this study the breakdown of these costs are provided in sections 4.4.1 and 0.

4.4.1 Fixed costs

Fixed costs include:

- direct labour
- administration and support labour
- maintenance

4.4.1.1 Direct labour

Direct labour was calculated on a 70000\$/year/operator basis. The number of operators was adopted from the IEA report number 2005/09 to be 144 units for the ASC oxy-coal case and 112 units for the ASC air-fired coal power plant case. However it should be mentioned that the number of operators reported for the oxy-fuel plant (136 units) was believed to be an underestimation, since the design did not include a flue gas desulfurization unit (FGD), thus an additional 8 operators for the FGD unit as reported by the same study was implemented [14].

4.4.1.2 Administrative and support labour

This component of the fixed costs is commonly implemented as a percentage of total labour costs. In this study as commonly employed in literature the percentage implemented is 30%.

4.4.1.3 Maintenance

Some components of maintenance could potentially fall under variable costs however as suggested by IEA GHG 2006/01 report when detailed breakdown of process requirements along with historical data are not available for certain processes, maintenance costs are better represented exclusively as fixed costs. The same study suggests the breakdown for maintenance costs as a percentage of installed costs for different process areas which has been adopted in this study as well [19]. A summary of maintenance costs is provided in Table 10.

Table 10: Maintenance cost summary

		ASC oxy-coal with CO ₂ capture	Air fired ASC without CO ₂ capture
Process area/s	Maintenance cost as % of installed costs	Maintenance costs (million US \$/year)	Maintenance costs (million US \$/year)
Coal/sorbent preparation, ash & spent sorbent handling, boiler and accessories, flue gas cleanup, balance of plant	4%	\$ 22.3	\$ 17.4
CO ₂ capture & compression unit and air separation unit	2.5%	\$ 6.6	\$ 0
Utilities & off sites	1.7%	\$ 2.5	\$ 2.0

4.4.2 Variable costs

Variable costs include costs of consumables and wastes produced by any of the plant's comprising processes. It should be noted that fuel cost is commonly included as a component of the variable costs; however, in the financial model employed this cost is added as a separate entity. As for the consumables and waste disposal costs the numbers provided by the department of energy's report were employed to estimate variable costs [16]. The numbers expressed on a \$/MWhr_{net} were employed however it should be noted that since

there was no NO_x control process included in this study and the sulphur loading was approximately a quarter of that of the DOE's study these costs were respectively eliminated and scaled to match the process at hand for the ASC oxy-fuel case. Furthermore, consumables included makeup and cooling water, and any chemicals or sorbents employed. Since a water balance was not provided for this study and since the process configuration of the study at hand was similar to that of the department of energy's, the \$/MWhr_{net} values from the department of energy's study were deemed appropriate for use in this study. A summary of variable and O&M costs can be seen in the Table 11.

Table 11: Operations and maintenance cost summary

	ASC oxy-coal with CO₂ capture	Air fired ASC without CO₂ capture
Component	Costs (million US \$/year)	Costs (million US \$/year)
Direct labour	\$ 10.1	\$ 7.8
Administration and support labour	\$ 3.0	\$ 2.4
Maintenance	\$ 31.4	\$ 19.4
Total fixed O&M	\$ 44.5	\$ 29.6
SCR catalyst replacement	\$ 0	\$ 0.5
Consumables (excluding fuel)	\$ 3.6	\$ 2.9
Waste disposal	\$ 2.0	\$ 1.4
Total variable O&M	\$ 5.5	\$ 4.8
Total O&M costs	\$ 50.0	\$ 34.4

4.4.3 Comparison of O&M costs

In order to come up with comparable values it was assumed that for a full scale power plant the direct labour is not necessarily a function of power plant size rather it is more a function of the comprising process areas. A confirmation of this assumption is the detailed direct labour unit numbers provided by the two IEA reports 2005/09 and 2006/01 which were 136 and 122 for a 532 and a 741 MWe_{net} power plants respectively [14][19]. It was observed that the number of labour units considered for the smaller power plant was even more than the number of labour units for the larger power plant although the processes were similar in nature. Thus it was assumed that this number firstly requires a detailed analysis and availability of historical data, and secondly it is not really a function of the power plant size for full scale operations. Consequently for comparison purposes this number was not meddled with. As for maintenance costs the same approach as IEA report 2006/01 was employed and it should be noted that for all three studies an escalation rate of 2% was used for updating all costs [19]. It should be noted that the cost of fuel was eliminated from the costs in every case. A summary of these updated and adjusted costs can be seen in Figure 21:

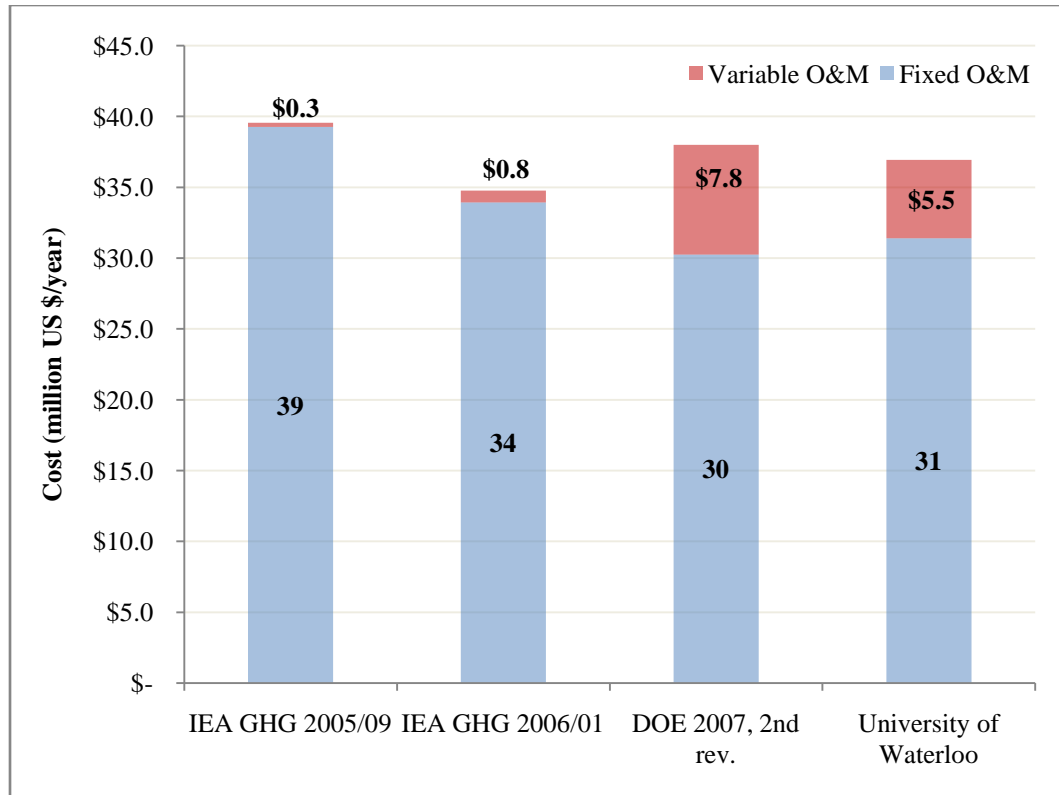


Figure 21: Comparison of O&M costs for a 380.5 MWnet ASC oxy-coal plant with CO₂ capture

A large variation is observed in the variable and fixed O&M costs since different breakdowns are employed for fixed and variable costs. The study conducted by the DOE considers maintenance material costs to be a part of variable costs whereas for the other two reference studies and for this study maintenance costs are exclusively reported as fixed costs.

Also, it is observed that although similar breakdowns are employed for this study and the two IEA GHG reports, there is a large difference observed in the variable costs reported for the ASC oxy-fuel case. This is due to the absence of an FGD unit in the two aforementioned references and thus the cost of sorbent is excluded which is one of the major contributors to the variable costs.

It should also be pointed out that the approach assumed by the University of Waterloo although a combination of various assumptions from different sources is a rather conservative approach as can be observed in the above graphs and explanations.

4.5 Summary of costs

A summary of the capital and O&M costs along with the calculated LCOE both for the with and without capture cases and the corresponding LACOC is provided in Table 12. It should be however noted that since large variations in cost indices was observed over the course of last year, an average of the past twelve months' CEPCI was employed, to update all scaled estimates. Furthermore O&M and fuel costs are escalated by an escalation factor of 2% per year over the life of plant in coming up with the presented costs.

Table 12: Summary of costs

	ASC oxy-coal with CO₂ capture	Air fired ASC without CO₂ capture
Estimated parameter	Costs (million US \$/year)	Costs (million US \$/year)
Total capital cost required (\$)*	\$ 1,446.6	\$ 857.4
Operation & Maintenance cost minus fuel (\$/year for first year)**	\$ 50.0	\$ 34.4
Fuel cost (\$/year for first year)**	\$ 45.9	\$ 40.1
Levelized cost of electricity(\$/MWhr _{net})	\$ 130.8	\$ 83.3
Levelized CO ₂ avoidance cost (\$/tonne)	\$ 63.5	N/A

5 DESIGN CONSIDERATIONS

5.1 CO₂ compression unit design considerations

5.1.1 Materials of construction

For all equipment prior to the drying unit, stainless steel 316L is employed which is commonly employed in sulphuric acid plants and is one of the more corrosion resistant of stainless steels. After the molecular sieve drying unit all process components were constructed from stainless steel 304 rather than carbon steel to provide further corrosion resistance.

5.1.2 Heat exchanger design

Heat exchangers were one of the most expensive unit operations employed in the CCU. Due to the corrosive nature of the gases, the gases in all heat exchangers in the CCU were directed through the tubes to reduce the cost of shell (being made of corrosion resistant SS 316L).

5.1.2.1 Cooling water

In all of the CCU coolers the assumption was that untreated cooling water at 25°C was employed thus considerations were implemented for ease of cleaning of heat exchangers. Also the maximum outlet water temperature was kept under 35°C due to the reverse solubility of minerals present in untreated water. Thus, fouling is kept under control by not exceeding 35°C. The TEMA (tubular exchanger manufacturing association) design CFU was employed in order to provide removable bundle capabilities for shell side cleaning. The tube pitch was also changed from a 30° triangular to 90° square and the pitch was increased to provide more than 6mm of cleaning space between tubes.

5.1.2.2 Further design considerations

The weight of all heat exchanger bundles were kept under 20,000 kg with maximum shell diameters being less than 2m in order to remain within crane limitations commonly employed.

All designs were analyzed for vibration issues and all issues were resolved by implementing sufficient tube supports for designs and changing baffle arrangements to no tubes in window arrangements single segmental from the common single segmental baffles; although this results in larger shell diameters it was deemed necessary in eliminating any vibration issues possible.

5.1.2.3 Overdesign factor

An overdesign factor of 5% was employed for all heat exchangers which is an industry accepted limit for large heat exchangers. This is considering the fact that proper fouling parameters were implemented both for untreated cooling water and acid gas fouling resistance values commonly accepted in industry.

5.1.3 Compressor design

All compressors were made out of stainless steel 304 although there is a small chance of any condensation of acid gases just to provide more corrosion protection. All compressors were upsized to the next available off the shelf size (according to Icarus). Polytropic efficiency of compressors were estimated based on the volumetric input of flue gas in order to provide more accurate designs along with compression ratios being maintained below five.

5.2 Flue gas clean up

5.2.1 Materials of construction

As for materials of construction all heat exchangers dealing with flue gases were designed using SS 316L stainless steel, with cooling water heat exchanger sides being made of SS 304. Recycle fans were also assumed to be constructed from stainless steel as well.

5.2.2 Heat exchanger design

The heat exchanger designs employed was the TEMA AXL. This design was employed in order to minimize the pressure drop, although the heat transfer coefficient is reduced for the most part due to the cross current nature of the flow, and temperature crosses (cross of the outlet of hot stream and inlet of cold stream) present in the heat exchangers proposed. Due to the high cost of these heat exchanger units and possible concerns of high sulphur recycled flue gas streams two design schemes were analyzed. It should be noted however that typical tubular heat exchangers designed for gas/gas heat exchange processes in power plants are not commonly cylindrical shell and tube, thus the price of these units are believed to be a slight overestimation.

5.2.2.1 Case scenario 1

Assumes there are two heat recovery exchangers after ESP and the flue gas desulphurization (FGD) unit is placed after the recycle point. The first heat recovery unit is a gas/gas heat exchanger and in the second one the cooling medium is condensed steam employed for further cooling the flue gas.

5.2.2.2 Case scenario 2

Due to the high cost of gas/gas heat exchangers the case was analyzed where the FGD is placed prior to the recycle stream and since it is a wet FGD system this unit can take on the duty of one of the heat recovery units. Thus HR1 which is a gas/gas heat exchanger can be eliminated. This scenario would also improve the heat transfer coefficient of the cooling

medium (recycled flue gas) for the tubular air heater (TAH) placed after the boiler and prior to the ESP, since the recycled flue gas stream recycled back from the FGD outlet will not be employed as the cooling medium in the first heat recovery unit (HR1) as in case scenario 1. Thus the temperature of this cooling gas will be lower prior to being used as the cooling medium for the TAH, compared to its temperature in scenario 1. Also since in this case scenario all of the flue gas is treated with FGD the corrosion concerns case specific to oxy-fuel combustion are eliminated, thus further cost savings are possible when it comes to boiler materials of construction. However it is worth noting that the FGD is the single most expensive unit of the flue gas cleanup process for both case scenarios considered so its size reduction would typically result in major cost savings.

5.2.3 Overdesign factor

An overdesign factor of 5% was implemented for all units of the flue gas cleanup process.

6 SENSITIVITY ANALYSIS

Due to the case specific nature of power plants' designs and evaluations it is vital that the sensitivity of the reported costs be assessed in detail to suspected parameters of significance. Thus several case scenarios were developed illustrating the effect of the following parameters on the overall economics of this power project evaluation:

- Chemical Engineering Plant Cost Index (CEPCI)
- Fuel price
- Capacity factor
- Interest rate
- Debt ratio
- Construction duration
- Economic life of plant
- Power plant size

6.1 Chemical engineering plant cost index

Large deviations were observed in the chemical engineering plant cost index reported during the course of last year (May 2008 – April 2009). Thus, it was decided to employ an average of the last twelve months' reported indices by calculating the mean of the minimum and the maximum of the reported numbers. This average was employed as the plant cost index for updating all costs for the purposes of this report. Figure 22 illustrates the sensitivity of the LCOE with and without capture and CO₂ avoidance cost to the CEPCI. It can be observed that LCOE with and without capture vary from 121 to 140 \$/MWhr_{net} and 78 to 89 \$/MWhr_{net} respectively and the CO₂ avoidance cost varies from 58 to 69 \$/tonne depending on which month's CEPCI is employed as the reference month varying between 511.8 and 619.3 in the past twelve months. Due to the recent economic instability it was decided that both high and low indices reported during the course of last year can be ignored and this task

was fulfilled by averaging the minimum and maximum of the reported indices, which resulted in an average index of 565.55.

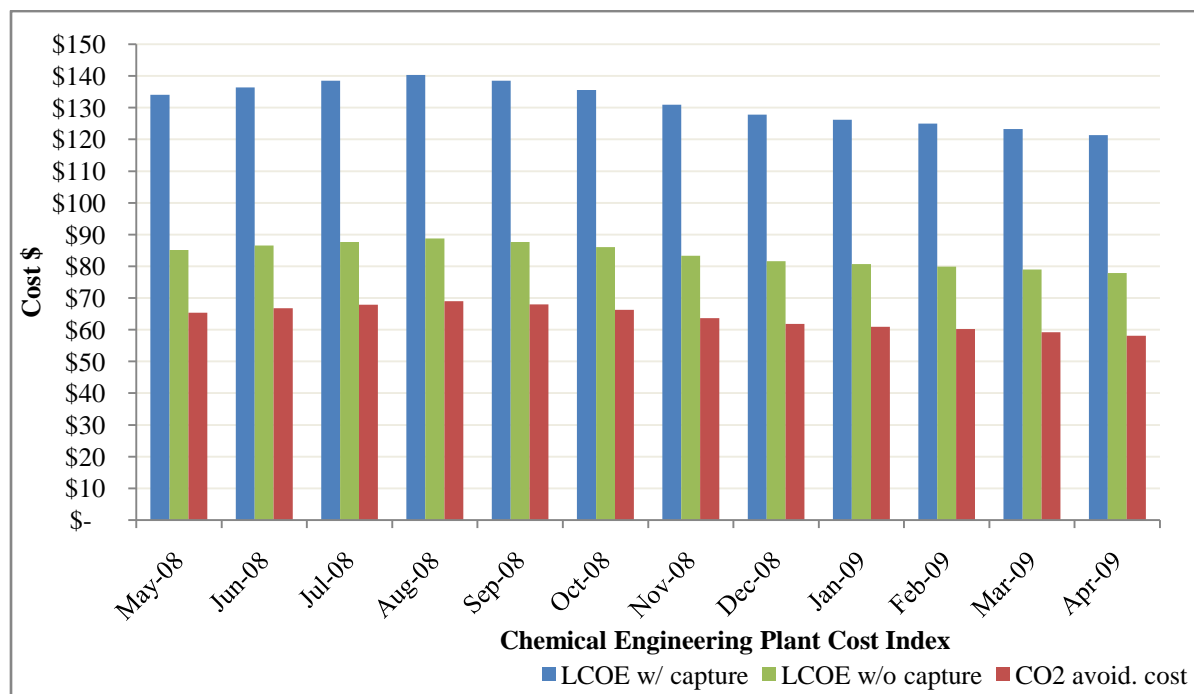


Figure 22: Sensitivity of LCOE with and without capture & LACOC to CEPCI

6.2 Fuel price

One of the influential parameters in the economics of fossil fuel power plants which typically experiences large fluctuations is the cost of fuel. Thus, it was deemed necessary to assess the sensitivity of the power plants' overall economics to the fuel price. It can be observed from Figure 23 for a (+/-) 50% change in price of lignite the LCOE with and without capture vary from 122 to 143 \$/MWh_{net} and 75 to 94 \$/MWh_{net} and the CO₂ avoidance cost varies from 62 and 66 \$/tonne. It should be noted that the default fuel price implemented is 23.41 \$/tonne.

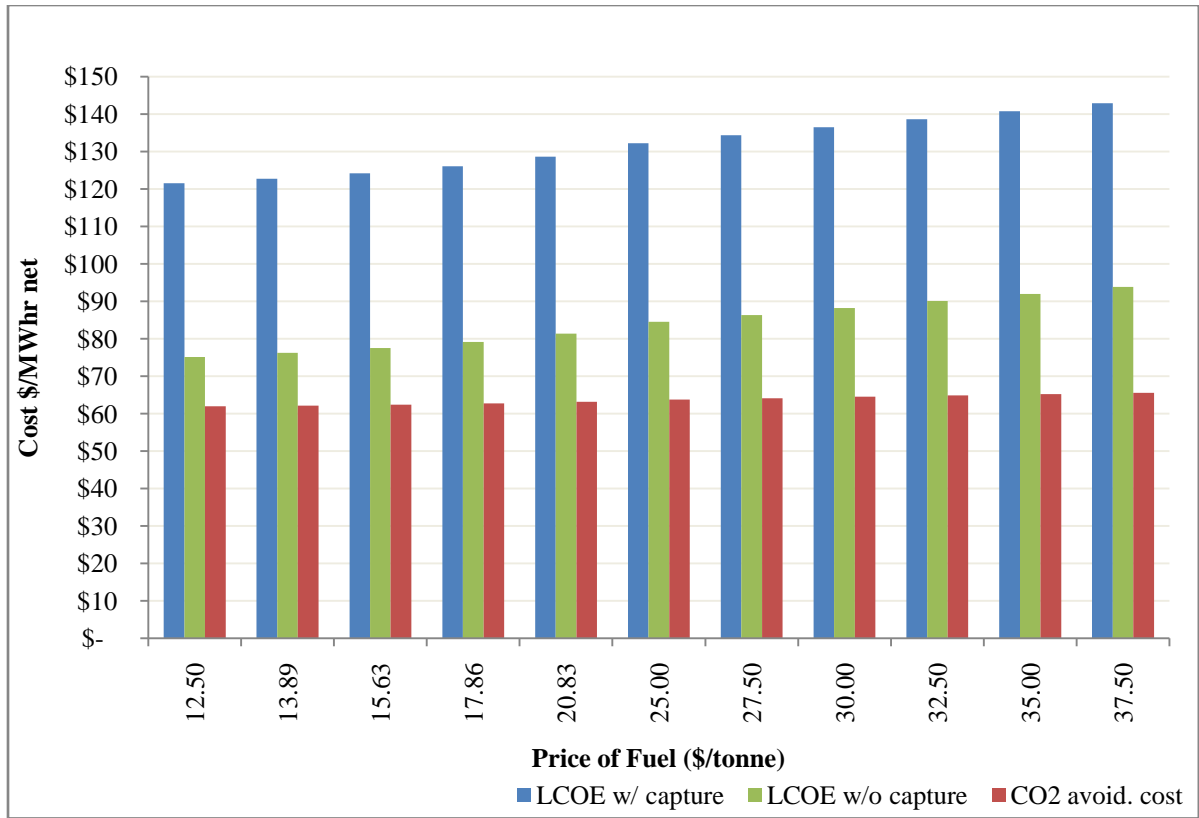


Figure 23: Sensitivity of LCOE with and without capture & LACOC to fuel price

6.3 Capacity factor

For an emerging technology one of the most important parameters of great concern is the capacity factor. With increased complexity and integration of a power plant operation reduced capacity factor is typically expected. Thus the sensitivity of the overall economics to variations in capacity factor was assessed and it was observed that the LCOE with and without capture vary from 119 to 145 \$/MWhr_{net} and 77 to 92 \$/MWhr_{net} respectively while the CO₂ avoidance cost varied between 57 and 72 \$/tonne when the capacity factor changes from 95% to 75% respectively. Figure 24 summarizes the aforementioned results. It should be noted that the default capacity factor employed in the evaluations was 85%.

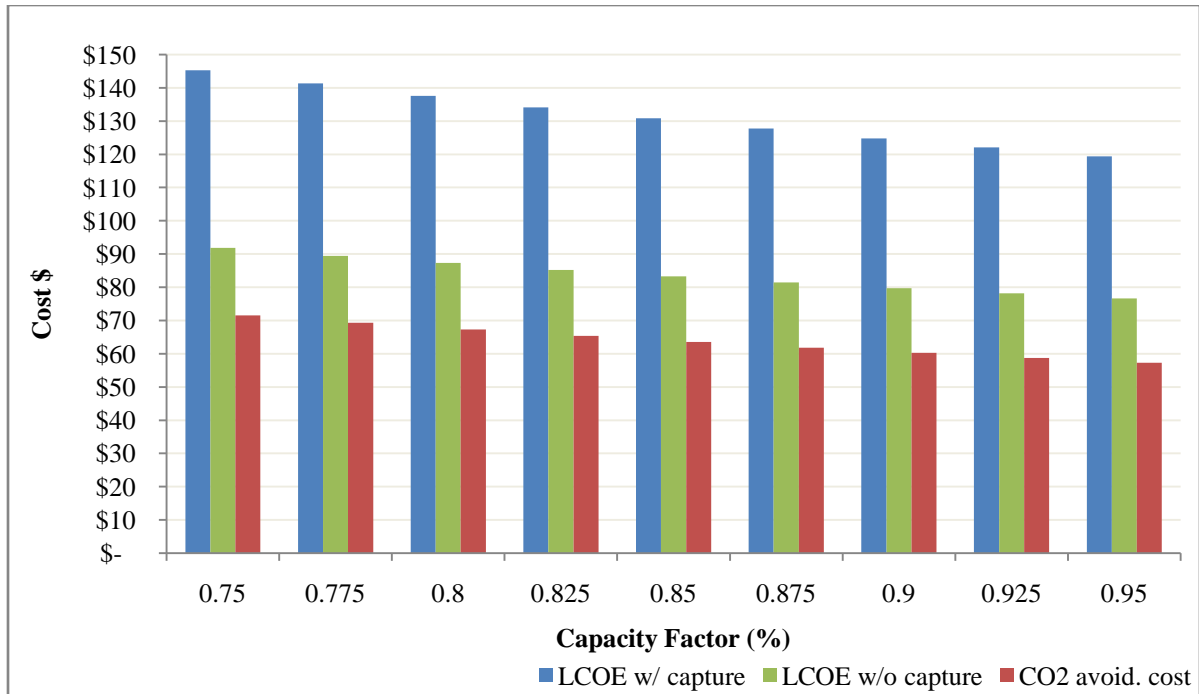


Figure 24: Sensitivity of LCOE with and without capture & LACOC to capacity factor

6.4 Interest rate

Interest rate is a parameter that directly affects the overall economics of a power project and is a function of the global economics situation as well as the nature of the project at hand. The sensitivity of the overall economics was assessed for a range of variation of five percent in the debt interest rate. The LCOE with and without capture were estimated to vary from 126 to 138 \$/MWh_{net} and 81 to 87 \$/MWh_{net} respectively and the CO₂ avoidance cost varied from 61 to 68 \$/tonne as the interest is increased from 5% to 10%. Figure 25 illustrates the impact of interest rate on the power project's overall economics. It should be noted that the default interest rate employed for the evaluations in this study was 7%.

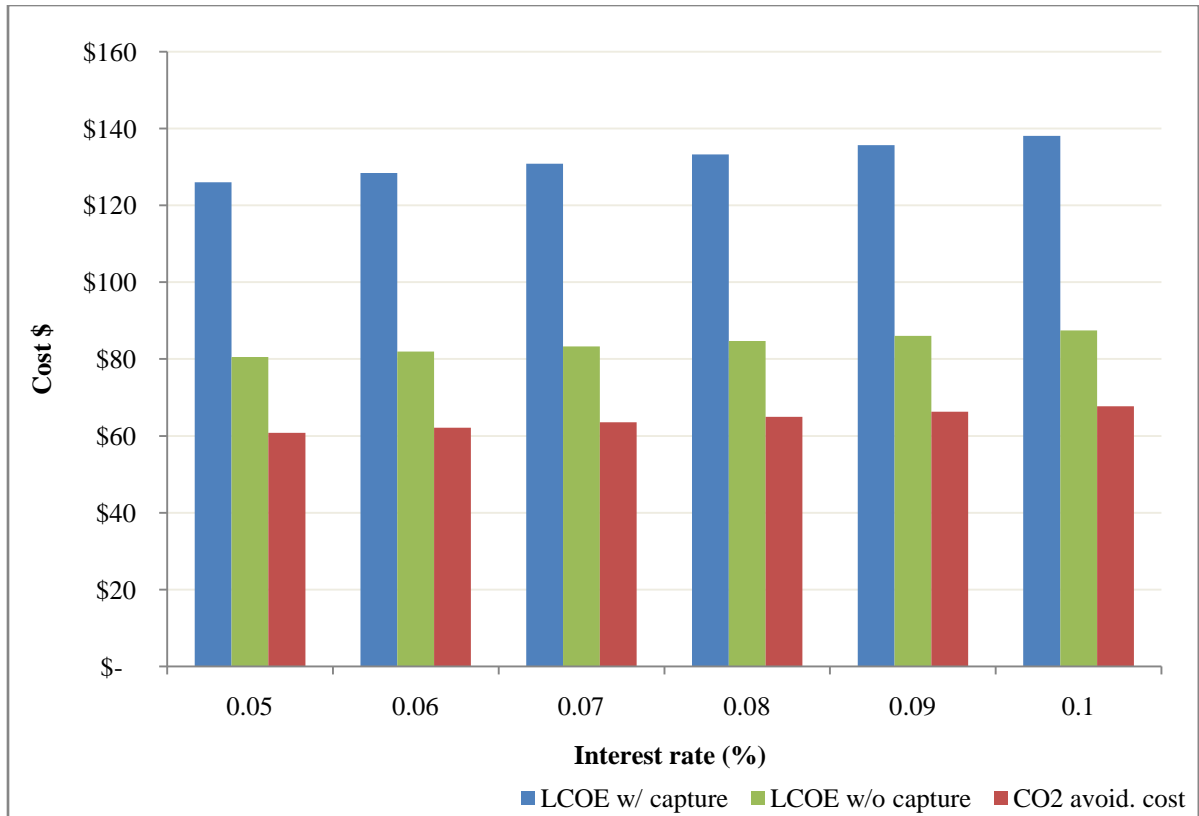


Figure 25: Sensitivity of LCOE with and without capture & LACOC to interest rate

6.5 Debt ratio

Debt ratio presents the fraction of invested capital that is provided from debt investment. Depending on how risky a project is this fraction changes since the balance is commonly provided by shareholders. When an investor is unsure of the nature of a project and the probability of its success, his/her tendency to invest might change. Since the expected returns on equity are typically higher than the debt return rates, changes in debt ratio are expected to affect the overall economics of the process. Consequently the sensitivity of the economics to a range of 0.4 to 0.6 has been assessed. Figure 26 indicates that when the debt ratio is changed from 0.4 to 0.6 the LCOE with and without capture were decreased from 138 to 124 \$/MWh_{net} and 87 to 79 \$/MWh_{net} respectively and the CO₂ avoidance cost decreased from 68 to 59 \$/tonne. It should be noted that the default debt ratio employed is 0.5.

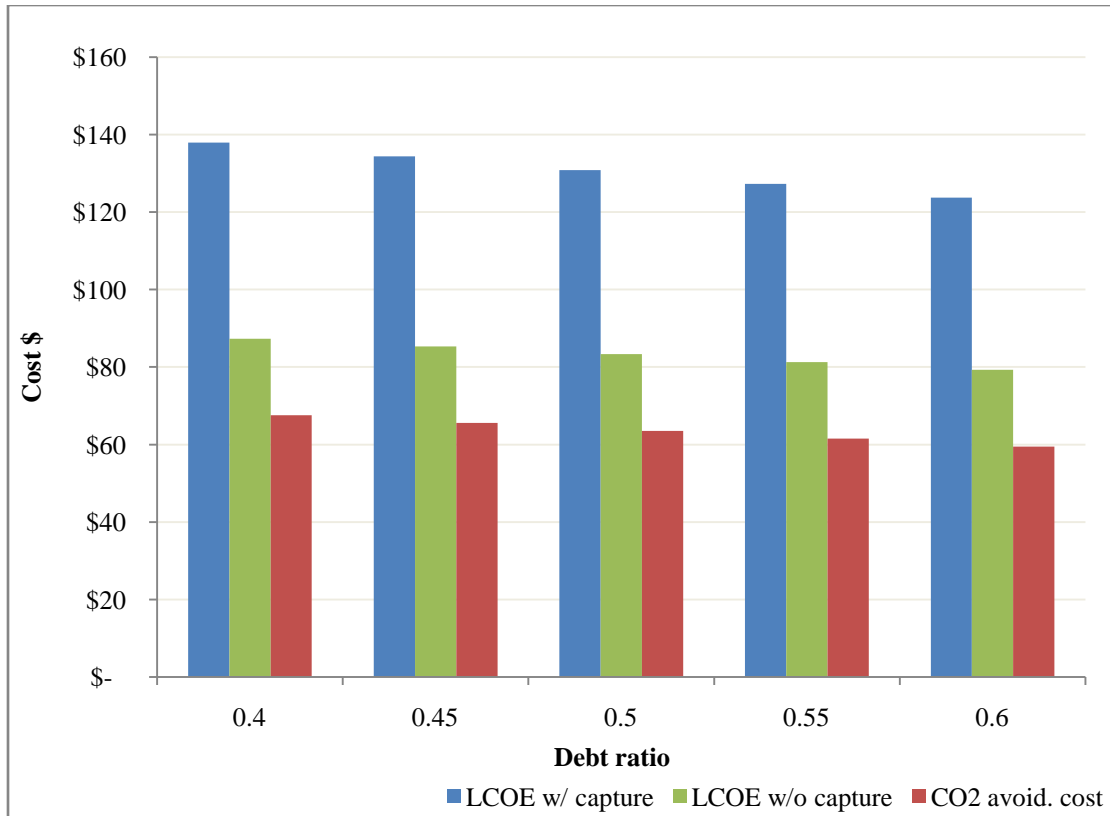


Figure 26: Sensitivity of LCOE with and without capture & LACOC to debt ratio

6.6 Construction duration

Duration of construction differs depending on the complexity and location of the power plant. For a coal plant the expected durations of construction varies from 3 to 6 years. During the construction period since a 100% debt financing is commonly employed at a higher interest rate than the long term financing of the power plant, the length of construction is expected to impact the overall economics. A sensitivity analysis is developed for construction durations of 3 to 6 years; the LCOE with and without capture respectively changed from 128 to 138 \$/MWh_{net} and from 81 to 87 \$/MWh_{net} while the CO₂ avoidance cost increased from 62 to 68 \$/tonne (Figure 27). It should be noted that no escalation in prices was accounted for during the construction period, inclusion of these escalation parameters are expected to further amplify the increase in both LCOE and CO₂ avoidance cost with increases in construction duration. It should be noted that the default construction

duration was 4 years for the ASC oxy-fuel case and 3 years for the ASC air-fired coal case in this study.

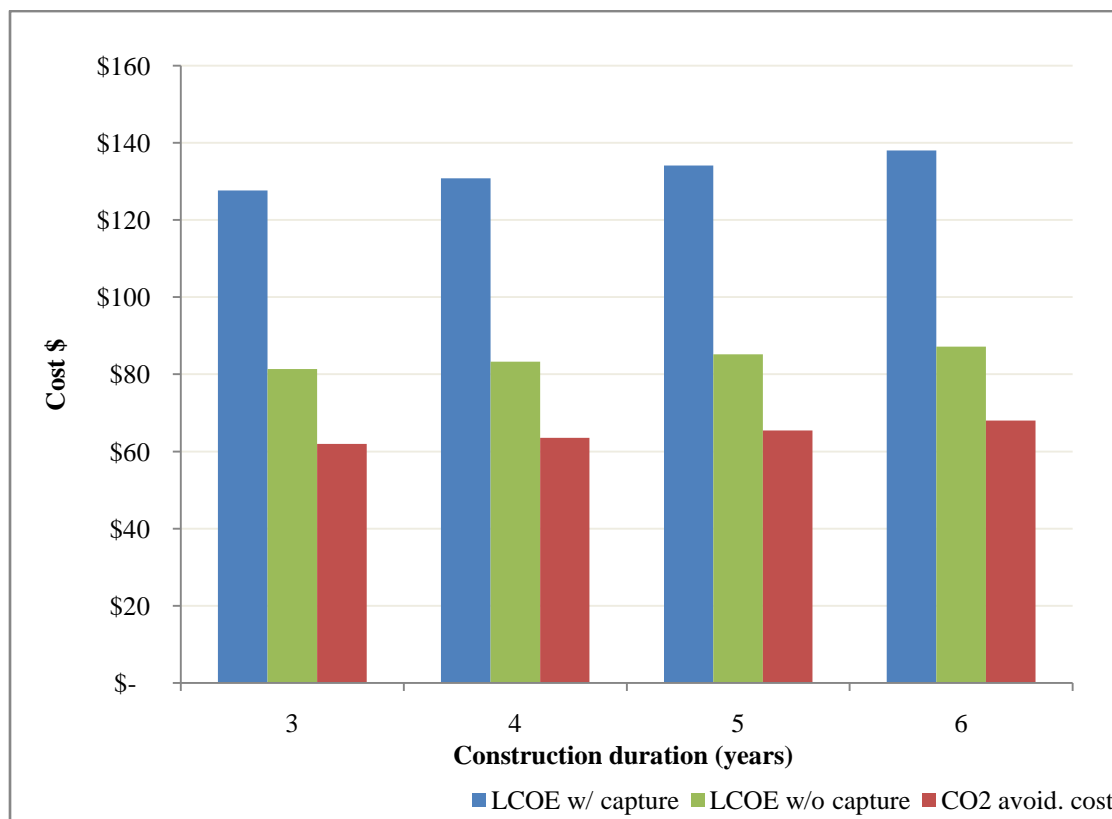


Figure 27: Sensitivity of LCOE with and without capture & LACOC to construction duration

6.7 Economic life of plant

In Canada there is a large debate as to how long the economic life of an oxy-coal plant should be assumed. Most recent sources in the literature assume numbers in the range of 20-30 years. However it is observed that typical Canadian coal plants are operated for over 40 years. Thus it was deemed necessary to assess the effect of the economic life on the overall economics. Figure 28 indicates that for an increase from 20 to 45 years of the economic life of plant the LCOE with and without capture decreased from 139 to 127 \$/MWhr_{net} and 88 to 81 \$/MWhr_{net} respectively while the CO₂ avoidance cost decreased from 68 to 61\$/tonne. It should be noted that the default plant economic life employed in this study is 30 years.

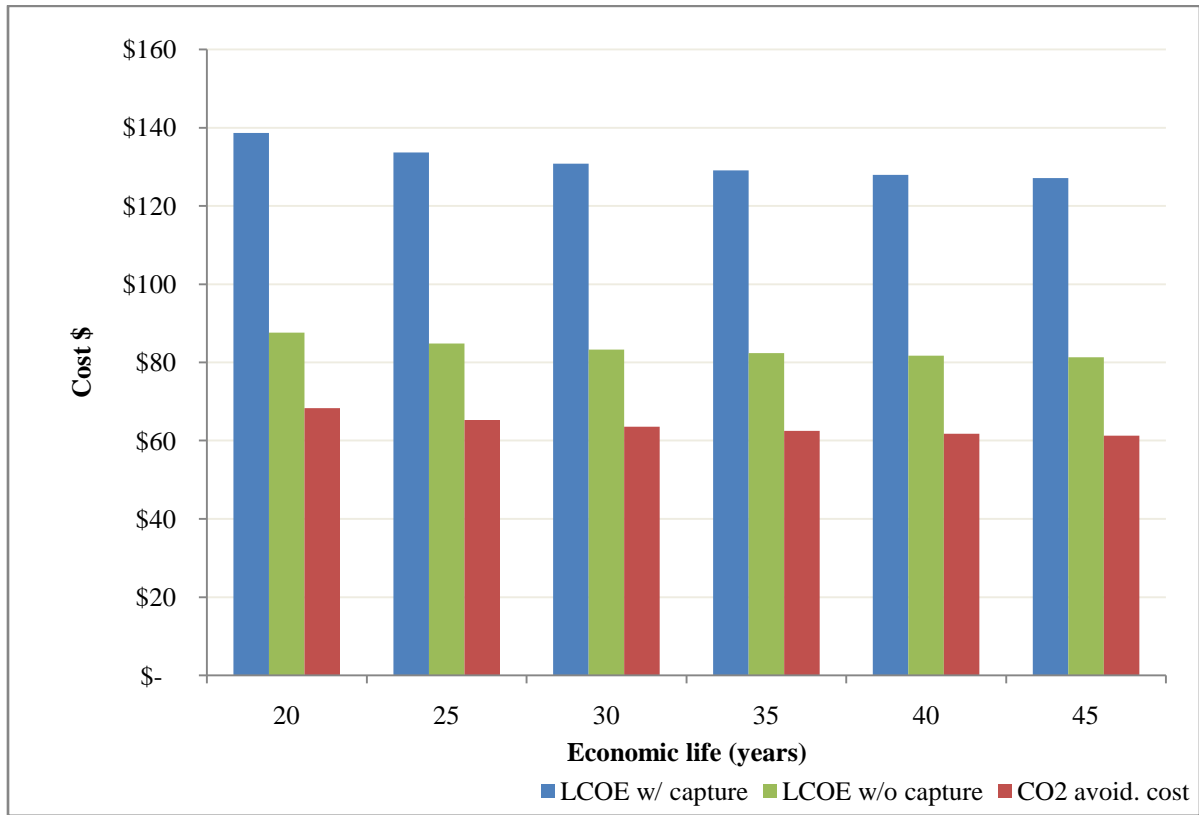


Figure 28: Sensitivity of LCOE with and without capture & LACOC to economic life of plant

6.8 Power plant size

Since the model has the capability of developing costs for power plants in the range of 200-700+ MWhr_{net}, it was interesting to observe the effect of change in plant size on the overall economics. It was observed that for an increase in plant size from 200 to 700 MWhr_{net} the LCOE with and without capture varied from 161 to 117 \$/MWhr_{net} and 101 to 78 \$/MWhr_{net} respectively while the CO₂ avoidance cost decreased from 80 to 52 \$/tonne as summarized in Figure 29.

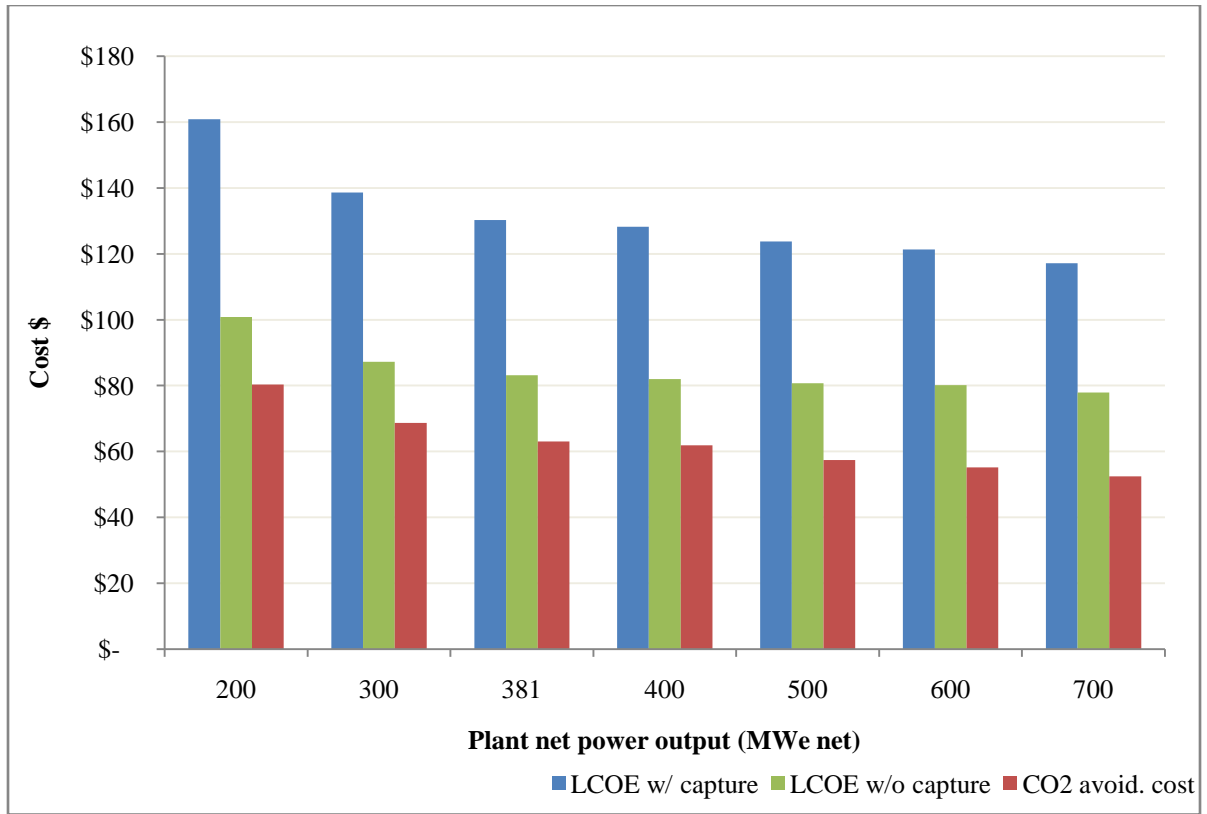


Figure 29: Sensitivity of LCOE with and without capture & LACOC to plant size

6.9 Sensitivity analysis summary

Two summary graphs illustrating the relative effect of the parameters under study on the overall economics of the process is prepared since it is interesting to identify which parameters are the most influential of all, Figure 30 and Figure 31. In these summary graphs the x-axis represents the percentage change in the parameter under study relative to the default value employed for the reported costs and the y-axis represents the resulting percentage change in the LCOE with CO₂ capture in Figure 30 and the resulting percentage change in the LACOC in Figure 31. It can be observed that all parameters significantly affect the overall economics. However, it can be noted that the ordered parameters from most influential to least influential on the LCOE with CO₂ capture are the chemical engineering plant cost index (CEPCI), capacity factor, size of power plant, debt ratio, economic life of plant, fuel price, interest rate, and construction duration. Similarly the same

ordering prepared for the LACOC is chemical engineering plant cost index (CEPCI), capacity factor, size of power plant, debt ratio, economic life of plant, interest rate, construction duration, and fuel price. It should be noted that there might be a slight overlap in the above mentioned orderings due to the non-linear behaviour of the over-all economic to changes in certain parameters, however this overlap is not very significant.

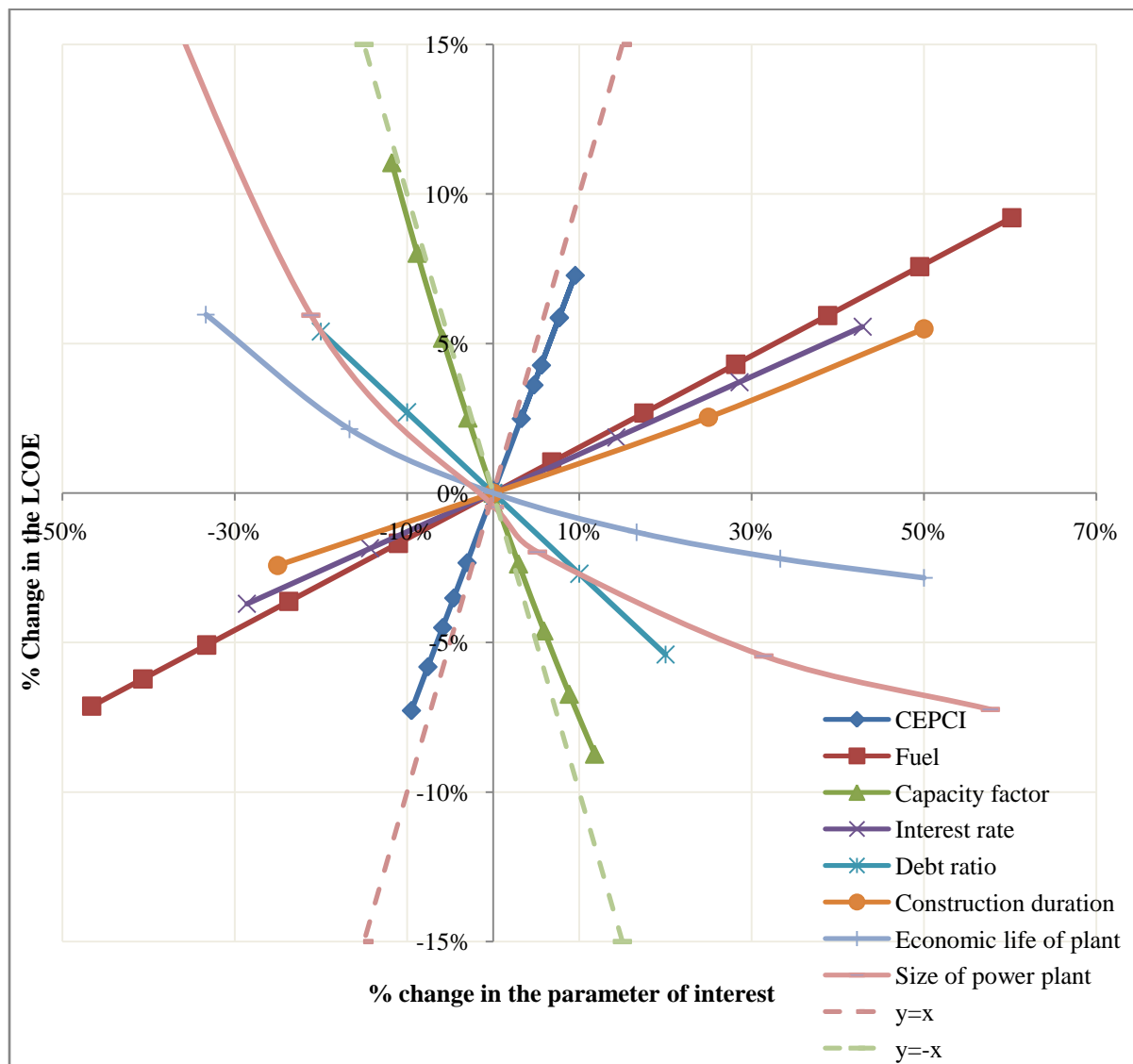


Figure 30: Summary of sensitivities of LCOE with CO₂ capture to parameters of consideration

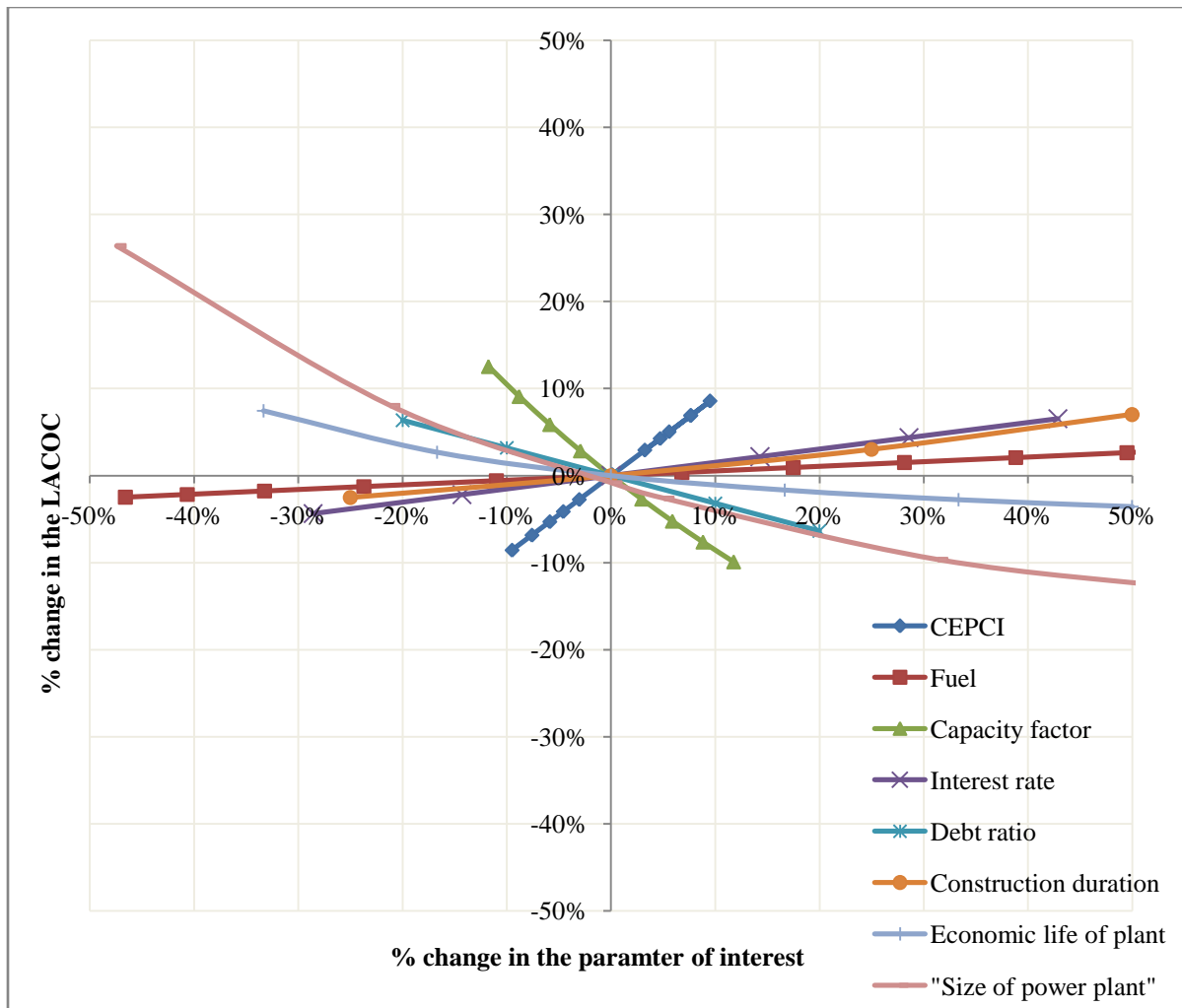


Figure 31: Summary of sensitivities of LACOC to parameters of consideration

7 CONCLUSIONS & RECOMMENDATIONS

With the current state of knowledge one of the most economical power production technologies with capture capabilities appear to be oxy-fuel combustion along with Amine absorption technology when looking at cost comparisons available in the literature for greenfield power plants.

A detailed financial model was developed in this study and levelized costs of electricity of 131 \$/MWhr_{net} and 83 \$/MWhr_{net} for a 380.5 MW_{net} ASC oxy-coal with CO₂ capture and a 380.5 ASC air-fired coal power plant without CO₂ capture capabilities respectively were estimated. Also a CO₂ avoidance cost of 64 \$/tonne was calculated.

The sensitivity of the overall process economics was assessed against variations in several parameters and it was observed that the levelized cost of electricity with CO₂ capture was sensitive to the following parameters in the following descending order:

- CEPCI
- capacity factor
- size of power plant
- debt ratio
- economic life of plant
- fuel price
- interest rate
- construction duration

In parallel the levelized cost of CO₂ avoidance was found to be sensitive to the following parameters in the following descending order:

- CEPCI
- capacity factor

- size of power plant
- debt ratio
- economic life of plant
- interest rate
- construction duration
- fuel price

It should be noted that in regards to a certain technology's economics the technology deployment curve for coal should be considered, since it sheds some light on what is to be expected for a technology according to the current state of knowledge.

More case specific information would assist in confidently reporting higher accuracy levels when employing costing software thus tightening the accuracy ranges reported on the final costs. For areas where costing software is not available vendor confirmations and quotations are much recommended to ensure the validity of costs.

The determination of emission intensities of the power plants both with and without capture capability is much recommended. These two parameters significantly affect the avoidance costs determined for the technology.

Including the water cycle would also be of great importance since the effect of more complex heat integration schemes can be better realized in the determination of variable costs and also since water shortage is a global environmental concern.

Due to large accuracy bounds reported for the boiler, balance of plant and air separation unit and the rapid process improvements expected in air separation technologies obtaining detailed vendor quotations for any of these units would bring about large improvements in the overall reported cost accuracies.

Detailed FGD model developments would also assist in obtaining more accurate costs for the different case scenarios evaluated since it is one of the major costs included in the flue gas cleanup. Furthermore obtaining quotations for the gas/gas heat recovery units which are not cylindrical shell and tube heat exchangers but were simulated as such is recommended.

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APPENDICES

Appendix 1: vendor quotations

Aluminum brazed heat exchangers



BUDGET QUOTE

Chart Energy & Chemicals
2191 Ward Avenue
La Crosse, WI 54601 USA
Phone: 608.787.3333 Fax: 608.787.2141
www.chart-ind.com

TO: Canmet Energy **DATE:** APRIL 16, 2009
ATT: Ahmed Shafeen **PAGES:** 2
FROM: Dave Scott,
Ramco Energy Products Ltd.

SUBJECT: CO2 RECOVERY FULL SCALE PLANT

Chart Energy & Chemical's is pleased to present our budgetary proposal for the following brazed aluminum heat exchangers for the subject project.

I. EQUIPMENT DESCRIPTION AND PRICING

<u>Item #</u>	<u>Spec Sheet</u>	<u>Core Size (in)</u>	<u>No. of Cores</u>	<u>Unit Price (Ex-Works)</u>
E-102 WHE	JEH240	42" x 64" x 102"	4	US \$1,925,000
E-103 CHE	JEH239	48" x 114.4" x 96"	1	US \$455,000

Pricing includes:

- Standard Side support angles for mounting purposes
- 6061-T6 RFWN flanges on all connections
- Manifolding for E-102 (4 core) battery assembly, single connections per stream

II. TECHNICAL COMMENTS

- All heat exchangers are designed and constructed to the ASME Code Section VIII Div. 1.
- All equipment will be manufactured according to Chart's standards, as well as the standards set forth by ALPEMA. All heat exchangers will receive hydrostatic (1.3 x design pressure) or pneumatic testing (1.10 x design pressure) as required by Chart's standards and the current ASME Code. No special testing has been included other than that required by ASME code. Please advise if any special testing is required.
- Streams to be pressurized with dry air or N2 (to 15 psig) for shipment.
- Perforated plates have been included on E-102 stream 43-44 and E-103 stream 38-39 to ensure proper 2-phase flow distribution into the exchanger.
- Rating cases will be analyzed upon request for firm quote.



BUDGET QUOTE

Chart Energy & Chemicals

2191 Ward Avenue
La Crosse, WI 54601 USA
Phone: 608.787.3333 Fax: 608.787.2141
www.chart-ind.com

- F. E-102 design is pressure drop limited and optimization with greater allowable pressure drop, if available, can possibly reduce the number of cores

III. COMMERCIAL COMMENTS

- A. Price includes domestic shipping skid.
- B. Price does not include duty or taxes.
- C. Pricing is quoted Ex-works Chart facility.
- D. Payment terms are as follows for any resulting purchase order:
 - 20% at Chart drawing submittal
 - 50% at Chart midpoint of cycle
 - 30% at Chart offer to ship
- E. Chart standard terms and conditions of sale shall apply.
- F. The above price offered is budgetary only and not valid for purchase.

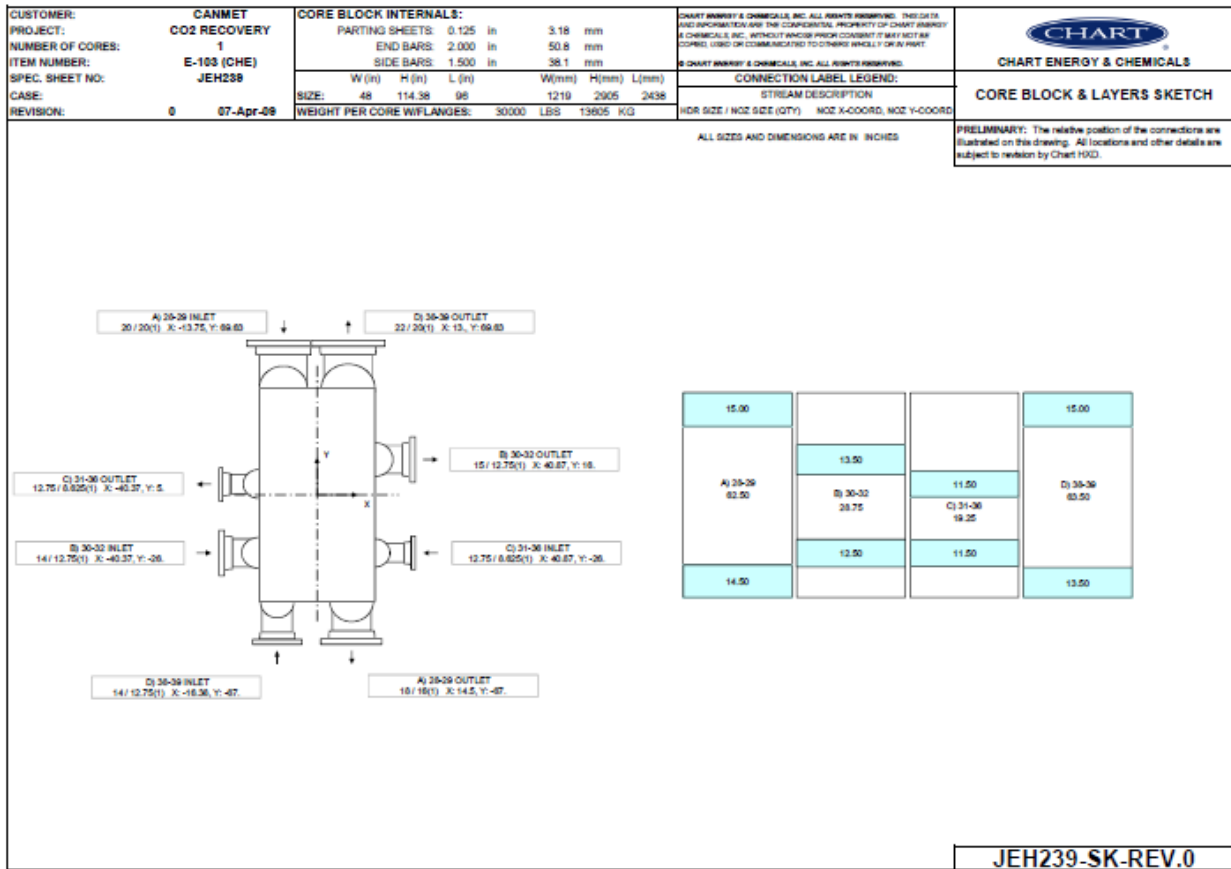
IV. SHIPPING CYCLE

The estimated lead time is 44 weeks for E-103 and 54 weeks for E-102. Chart typically can provide drawing submittals approximately (8-10) weeks after contract award. Delivery times can vary as our workload fluctuates, so we would ask that delivery times be re-confirmed prior to an award.

We trust that this proposal meets your expectations in full. If you have any questions or concerns, please feel free to call us at 403-252-3336.

Best Regards,

Dave Scott,
Ramco Energy Products Ltd.
Ph: 403-252-3336
Cell: 403-818-7270



JEH239-SK-REV.0



Molecular sieve CO₂ dryer



P.O. Box 2500, Richmond, KY 40476-2602

PH: 859-624-2091 / FAX: 859-623-2436

Natural Resources
2009

Date: May 11,

137-JB1

Quotation NO: 09-

ATTN: Ashkan

PH: 613-996-5371

Email: abeigzad@nrcan.gc.ca

abeigzad@yahoo.ca

REF:

APPLICATION: Drying 509,913 Kg/Hr of Flue Gas at 30 BAR and 40°C containing 600 Kg/Hr of water to a -40°C dewpoint.

3 B-80M Lectrodryer dual tower heat regenerated desiccant dryer for fully automatic operation.

Reactivation gas and system to be determined.

Typical features would include ASME code-stamped pressure vessels, initial charge of desiccant, stainless steel desiccant supports, 24" desiccant fill/manway and bottom manway, interconnecting piping arrangement with pneumatically operated 2-way valves, relief valves, pressure transmitter and inlet and outlet temperature sensors for each adsorber, pressure gages, and all required instrumentation and moisture transmitter wired to skid PLC control with operator interface display for fully automatic dryer operation and status indication, and remote DCS monitoring. Electricals classification has not been specified.

APPROVAL DRAWINGS: 10-12 weeks after order

SHIPMENT: First dryer 26 weeks after drawing approval

BUDGET PRICE: FCA Richmond, KY	(+/- 30%)	\$4,500,000/ Per Unit
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TOTAL BUDGET PRICE: FCA Richmond, KY	(+/- 30%)	\$13,500,000 / 3 Units
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Natural Resources

Quotation 09-137-JB1

Page 2 of 2

COMMENTS:

- 1) Above equipment quote based on the experience of Lectrodryer, but with no specifications.
- 2) Adsorber towers, platform, and upper piping are shipped loose requiring reassembly on-site. All insulation by others.
- 3) Dimensions approximately 48 x 27 x 35 feet H.
- 4) Process pressure drop < 10 PSID.
- 5) Dryers quoted are double the size of the largest skids Lectrodryer has designed and sold. Quote should be considered adequate only for rough budget estimates.

* Quotation valid for 90 days

Best regards,

A handwritten signature in black ink, appearing to read 'John McPhearson', with a large, stylized initial 'J' and a horizontal line extending to the right.

John McPhearson
CEO
LECTRODRYER LLC

Attachments: B Bulletin